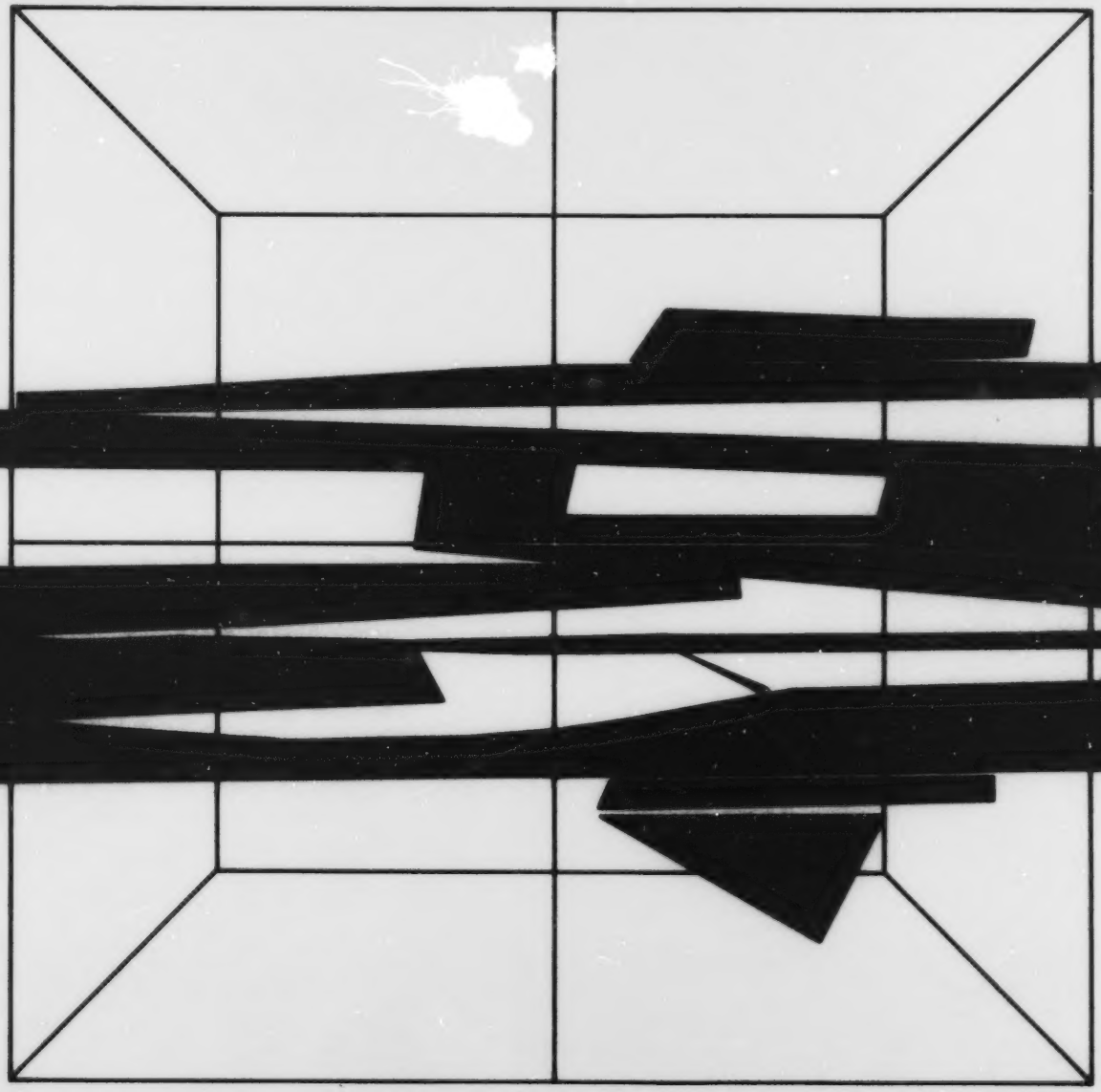
Water and Waste Management Practices for Coal-fuelled Power Generating Stations Equipped with Flue Gas Desulphurization

1885 1985

Report EPS 2/PG/1
February 1985



Environment Canada
Environmental Protection Service

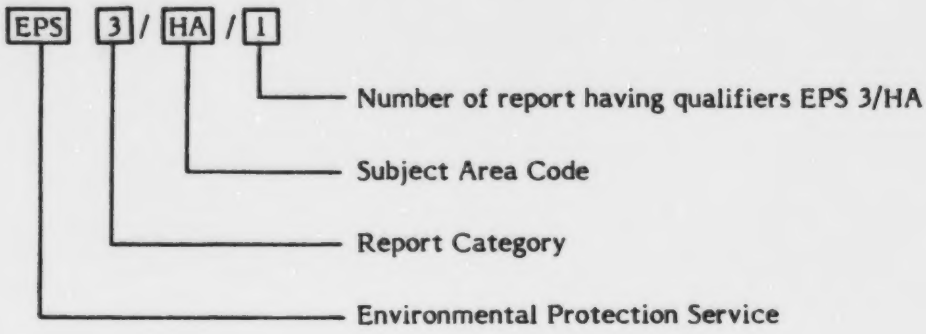
Environnement Canada
Service de la protection de l'environnement

Canada

85-3998/

ENVIRONMENTAL PROTECTION SERVICE REPORT SERIES

Sample Number:



Categories

1	Regulations/Guidelines/Codes of Practice
2	Public Consultation: Problem Assessment, Control Options
3	Research and Technology Development
4	Literature Reviews
5	Surveys
6	Social, Economic and Environmental Impact Assessments
7	Surveillance
8	Policy Proposals & Statements
9	Manuals

Subject Areas

AG	Agriculture
AP	Airborne Pollutants
CC	Commercial Chemicals
CP	Consumer Pollution
FP	Food Processing
HA	Hazardous Wastes
IC	Inorganic Chemicals
MA	Marine Pollution
MM	Mining & Mineral Processing
NR	Northern Regions
PF	Paper and Fibres
PG	Power Generation
PN	Petroleum and Natural Gas
SP	Oil & Chemical Spills
SRM	Standard Reference Methods
TS	Transportation Systems
UP	Urban Pollution

New subject areas and codes are introduced as they become necessary. A list of EPS reports may be obtained from the Publications Section, Environmental Protection Service, Environment Canada, Ottawa, Ontario, Canada, K1A 1C8.

National Parks Centennial
1885
1985
Centenaire des parcs nationaux

1885–1985:
100 years of Heritage Conservation
Canada's first national park was established in 1885 at Banff, Alberta. Today there are national parks and national historic parks in every province and territory. The National Parks Centennial is an occasion to renew our commitment to preserve examples of our heritage unimpaired for the benefit of all Canadians.

WATER AND WASTE MANAGEMENT PRACTICES FOR COAL-FUELLED POWER GENERATING STATIONS EQUIPPED WITH FLUE GAS DESULPHURIZATION

by

CH2M Hill Canada Ltd.

and

Dearborn Environmental Consulting Services Ltd.

for the

Industrial Programs Branch
Environmental Protection Service
Environment Canada

Report EPS 2/PG/1
February 1985

REVIEW NOTICE

This report has been reviewed by the Environmental Protection Programs Directorate, Environmental Protection Service, and approved for publication. Approval does not necessarily signify that the contents reflect the views and policies of the Environmental Protection Service. Mention of trade names or commercial products does not constitute recommendation or endorsement for use.

Any comments concerning the contents of this report should be directed to:

Mr. P.G. Finlay
Industrial Programs Branch
Environmental Protection Service
Environment Canada
Ottawa, Ontario
K1A 1C8

© Minister of Supply and Services Canada 1985
Cat. No. En 49-9/2-1E
ISBN 0-662-13742-6
BEAUREGARD PRESS LIMITED

ABSTRACT

This study investigates commercially demonstrated flue gas desulphurization (FGD) processes considered applicable to coals used for power generation in Canada (i.e., lime, limestone, dual alkali, alkaline fly ash, spray drying, Wellman-Lord, and seawater scrubbing). The water, wastewater, and solid waste implications of each process are evaluated and opportunities are identified to minimize the environmental impacts associated with the implementation of FGD technology in Canada.

Coal use in Canada is reviewed, and engineering and operating data are reported for several U.S. FGD facilities using coals considered representative of Canadian fuels. Characteristics are presented for samples of FGD sludge and wastewater collected during visits to these facilities.

Water consumption and liquid and solid waste production are compared for each of the FGD processes considered. FGD design options that could affect the quantity or characteristics of waste produced are identified, and alternative waste management strategies are reviewed, including waste utilization.

RÉSUMÉ

Sont étudiés divers procédés commercialement démontrés de désulfuration des gaz de combustion qui sont considérés comme applicables au Canada compte tenu des charbons utilisés pour la production d'électricité (c.-à-d. les procédés à la chaux, au calcaire, double alcali, d'épuration par les cendres volantes alcalines, de séchage par pulvérisation, Wellman-Lord et d'épuration à l'eau de mer). Les effets de chaque procédé sur la consommation d'eau et la production d'eaux usées et de déchets solides sont évalués, et des possibilités sont indiquées en vue de réduire les répercussions environnementales de l'application de techniques de désulfuration des gaz de combustion au Canada.

L'utilisation du charbon au Canada est passée en revue, et des données sont fournies sur la construction et le fonctionnement de plusieurs installations américaines de désulfuration des gaz de combustion où les charbons brûlés sont considérés comme représentatifs des combustibles canadiens. Des échantillons des boues et eaux résiduaires produites au cours de la désulfuration des gaz de combustion ont été prélevés lors de visites à ces installations, et leurs caractéristiques sont présentées.

Les chiffres de consommation d'eau et de production de déchets liquides et solides pour chacun des procédés examinés sont comparés. Les variantes de construction qui pourraient modifier la quantité ou les caractéristiques des déchets produits sont indiquées, et diverses stratégies possibles à propos des déchets sont examinées, y compris leur utilisation.

TABLE OF CONTENTS

		Page
ABSTRACT		iii
SUMMARY		xi
1	INTRODUCTION	1
1.1	Background	1
1.2	Objectives	1
1.3	Methodology	2
2	COAL-FUELLED GENERATING STATIONS IN CANADA	4
2.1	Introduction	4
2.2	Locations, Capacities and Coal Use at Existing Coal-Fuelled Generating Stations	4
2.3	Properties of Coal Used	5
2.4	Climatic Factors	11
3	FGD PROCESSES	14
3.1	Introduction	14
3.2	Limestone Scrubbing Process	16
3.3	Lime Scrubbing Process	19
3.4	Dual Alkali Scrubbing Process	22
3.5	Alkaline Fly Ash Scrubbing Process	24
3.6	Spray Dryer Scrubbing Process	26
3.7	Wellman-Lord Process	29
3.8	Other Processes	32
4	FGD-EQUIPPED STATIONS VISITED	39
5	WASTE CHARACTERIZATION	44
5.1	Introduction	44
5.2	Sample Procurement and Analytical Methods	44
5.3	Makeup Water and Wastewaters	48
5.4	FGD Sludges	52
5.5	Leaching Tests	57
5.6	Radiochemical Characteristics	65
5.7	Quality Control	65
5.8	Discussion and Conclusions	66
6	FGD PROCESS CONSIDERATIONS	68
6.1	Introduction	68
6.2	Material Balances	68
6.3	Process Design Alternatives	84
6.4	FGD System Reliability	99

		Page
7	WASTE MANAGEMENT CONSIDERATIONS	104
7.1	Introduction	104
7.2	FGD Waste Characteristics	104
7.3	Waste Processing and Disposal Alternatives	107
7.4	Waste Utilization	124
8	FGD SYSTEM COST	130
8.1	Sulphur Dioxide Removal Costs	130
8.2	Waste Management Costs	136
9	CONCLUSIONS AND RECOMMENDATIONS	139
9.1	Conclusions	139
9.2	Recommendations	142
REFERENCES		145
APPENDIX A	SITE VISIT REPORTS	153
APPENDIX B	DETAILED ANALYTICAL RESULTS	181

LIST OF FIGURES

Figure		Page
S-1	ALTERNATIVE FGD WASTE MANAGEMENT DESIGNS	xviii
1	LOCATION AND GENERATING CAPACITIES OF ONTARIO HYDRO'S COAL FIRED STEAM ELECTRIC STATIONS	5
2	LOCATIONS AND GENERATING CAPACITIES OF COAL-FUELLED GENERATING STATIONS IN THE MARITIMES AND WESTERN CANADA	7
3	LIMESTONE SCRUBBING PROCESS SCHEMATIC	17
4	SCHEMATIC OF JET BUBBLING REACTOR (JBR)	20
5	LIME SCRUBBING PROCESS SCHEMATIC	21
6	DUAL ALKALI SCRUBBING PROCESS SCHEMATIC	23
7	LIME/LIMESTONE ALKALINE FLY ASH SCRUBBING PROCESS SCHEMATIC	25
8	SPRAY DRYER SCRUBBING PROCESS SCHEMATIC	27
9	WELLMAN-LORD PROCESS SCHEMATIC	30
10	NORTHERN STATES POWER CO., SHERBURNE COUNTY GENERATING STATION SAMPLING LOCATIONS	45
11	MINNKOTA POWER COOPERATIVE INC., MILTON R. YOUNG GENERATING STATION SAMPLING LOCATIONS	46
12	MONTANA-DAKOTA UTILITIES CO., COYOTE STATION SAMPLING LOCATIONS	47
13	MILTON YOUNG GENERATING STATION FGD SLUDGE	54
14	SHERBURNE GENERATING STATION FGD SLUDGE	54
15	COYOTE GENERATING STATION DRY FGD SLUDGE	55
16	COYOTE GENERATING STATION CURED FGD SLUDGE	55
17	FLUE GAS SO_2 REMOVAL REQUIREMENTS AS A FUNCTION OF HEAT VALUE AND SULPHUR CONTENT OF COAL	70
18	MATERIAL BALANCE FOR LIMESTONE SCRUBBING PROCESS	73

Figure		Page
19	MATERIAL BALANCE FOR LIME-ALKALINE FLY ASH PROCESS	74
20	MATERIAL BALANCE FOR SPRAY DRYER PROCESS	75
21	MATERIAL BALANCE FOR SEAWATER SCRUBBER PROCESS	76
22	MATERIAL BALANCE FOR WELLMAN-LORD PROCESS	77
23	MATERIAL BALANCE FOR LIME SCRUBBING PROCESS	78
24	MATERIAL BALANCE FOR DUAL ALKALI SCRUBBING PROCESS	79
25	SINGLE-LOOP AND DOUBLE-LOOP SCRUBBING SYSTEMS	90
26	MEASURED EFFECT OF ORGANIC ACIDS ON SO_2 REMOVAL AT pH 5.0, L/G - 9 L/m^3	97
27	EFFECT OF 13.7 meq/L OF ORGANIC ACIDS ON LIMESTONE UTILIZATION AND SO_2 REMOVAL, L/G - 9 L/m^3	97
28	FGD SLUDGE DISPOSAL ALTERNATIVES	108
29	RELATIONSHIP BETWEEN SLUDGE VOLUME AND SOLIDS CONTENT	110
30	RELATIONSHIP BETWEEN FIXED FGD SLUDGE PERMEABILITY AND UNCONFINED COMPRESSIVE STRENGTH	113
31	POND UNDERDRAIN AND DECANT STRUCTURE	116
32	LAYOUT AND DESIGN OF A TYPICAL FGD SLUDGE LANDFILL	118
33	DESIGN AND CONSTRUCTION OF A TYPICAL GYPSUM STACK	119
34	CAPITAL INVESTMENT REQUIREMENTS FOR ALTERNATIVE FGD PROCESSES REPORTED BY NATO	132
35	ANNUAL REVENUE REQUIREMENTS FOR ALTERNATIVE FGD PROCESSES REPORTED BY NATO	132
36	CAPITAL INVESTMENT REQUIREMENTS FOR ALTERNATIVE FGD PROCESSES REPORTED BY PEDCo	134
37	ANNUAL REVENUE REQUIREMENTS FOR ALTERNATIVE FGD PROCESSES BY PEDCo	134

LIST OF TABLES

Table		Page
S-1	SUMMARY OF COAL TYPES AND FGD SYSTEMS EVALUATED	xii
S-2	WATER LOSSES FOR FGD PROCESSES	xiii
S-3	SOLID WASTE PRODUCTION RATES AND DISPOSAL VOLUMES	xv
1	GENERATING CAPACITIES AND COAL USE AT ONTARIO HYDRO COAL FIRED STEAM ELECTRIC STATIONS	6
2	GENERATING CAPACITIES AND COAL USE AT EASTERN AND WESTERN CANADIAN COAL FIRED GENERATING STATIONS	8
3	ANALYSIS OF COAL USED BY ONTATIO HYDRO IN 1981	9
4	PROXIMATE ANALYSIS OF COAL USED BY GENERATING STATIONS IN WESTERN CANADA AND THE MARITIMES	10
5	CLIMATIC DATA FOR ONTARIO HYDRO STATIONS	12
6	CLIMATIC DATA FOR SITES OF COAL-FUELLED GENERATING STATIONS IN THE MARITIMES AND WESTERN CANADA	13
7	DISTRIBUTON OF U.S. UTILITY FGD SYSTEMS BY PROCESS AS OF JANUARY 1982	15
8	SUMMARY OF STATIONS VISITED IN PHASE ONE	40
9	SUMMARY OF OPERATING CHARACTERISTICS OF STATIONS VISITED IN PHASE ONE	41
10	SUMMARY OF STATIONS VISITED IN PHASE TWO	42
11	SUMMARY OF OPERATING CHARACTERISTICS OF STATIONS VISITED IN PHASE TWO	43
12	SAMPLE COLLECTION SUMMARY	44
13	SAMPLE ANALYSES SUMMARY	49
14	COMPARISON OF ENVIRONMENT CANADA'S PROPOSED DISCHARGE CRITERIA WITH POTENTIAL WASTEWATERS FROM FGD SYSTEMS	51
15	ANALYSIS OF RAW AND LIME-TREATED SLUDGE RUNOFF SAMPLES FROM THE MILTON YOUNG GENERATING STATION	53
16	PHYSICAL PARAMETERS AND PRINCIPAL CONSTITUENTS OF FGD SOLID WASTES FROM SHERBURNE, MILTON YOUNG, AND COYOTE GENERATING STATIONS	56

Table		Page
17	PHYSICAL PARAMETERS AND PRINCIPAL CONSTITUENTS OF FGD SLUDGES FROM CANE RUN, DALLMAN, MARION, AND NEWTON GENERATING STATIONS	58
18	CHEMICAL COMPOSITION OF THE LEACHING MEDIA	59
19	RANGE OF CONCENTRATIONS OF CONSTITUENTS IN LABORATORY LEACHATES	60
20	CONSTITUENTS IN LABORATORY LEACHATES GROUPED BY RANGE OF CONCENTRATIONS	61
21	BEHAVIOUR OF FIVE METALS IN LEACHATES PRODUCED FROM MILTON YOUNG GENERATING STATION FGD SLUDGE	62
22	BEHAVIOUR OF FIVE METALS IN LEACHATES PRODUCED FROM COYOTE GENERATING STATION DRY (BAGHOUSE) FGD SOLIDS	62
23	BEHAVIOUR OF FIVE METALS IN LEACHATES PRODUCED FROM COYOTE GENERATING STATION CURED FGD WASTE	63
24	BEHAVIOUR OF FIVE METALS IN LEACHATES PRODUCED FROM SHERBURNE GENERATING STATION FGD SLUDGE	63
25	SUMMARY OF COAL TYPES AND FGD SYSTEMS EVALUATED	69
26	PROCESS DESIGN CRITERIA USED FOR FGD SYSTEM MATERIAL BALANCES	72
27	WATER REQUIREMENTS FOR FGD PROCESSES	81
28	SOLID WASTE PRODUCTION RATES AND DISPOSAL VOLUMES	83
29	AVERAGE ANNUAL PRECIPITATION INPUT TO SCRUBBER SLUDGE PONDS	94
30	PERCENT OF ALL PLANT OUTAGE TIME DIRECTLY CAUSED BY FGD COMPONENTS	100
31	TYPICAL SOLIDS CONCENTRATIONS OF FGD SLUDGES	111
32	CANADIAN GYPSUM PRODUCTION, TRADE, AND CONSUMPTION, 1979	127
33	SPECIFICATIONS FOR FGD BY-PRODUCT GYPSUM	128
34	LIMESTONE FGD SYSTEM COSTS	135
35	ESTIMATED CAPITAL INVESTMENT AND ANNUAL REVENUE REQUIREMENTS FOR ALTERNATIVE LIMESTONE FGD WASTE DISPOSAL SYSTEMS	137

SUMMARY

Introduction

With the promulgation of federal and provincial sulphur dioxide atmospheric emission guidelines, the Power Generation and Energy Division of Environment Canada commissioned this study to investigate the water, wastewater and solid waste implications of alternative flue gas desulphurization (FGD) systems. The study identifies opportunities to minimize the adverse environmental impacts associated with the implementation of FGD technology and investigates the reliability and cost of systems appropriate for use with Canadian coal.

Four coal types, considered representative of fuels used for thermal power generation in different Canadian regions, were examined. Material balances were prepared for different FGD systems applied to a hypothetical 500-MW(e) coal-fuelled station burning each of these four coals. Table S-1 summarizes the characteristics of the four fuels, the FGD systems examined for each, and the associated sulphur dioxide removals required to meet the National Guidelines for Atmospheric Emissions from New Thermal Power Generation Facilities.

Water Requirements and Wastewater Generation

Table S-2 summarizes the water losses in each of the FGD processes used with coal types considered in this study. Makeup water is added to the system to replace these losses, primarily for mist eliminator washing, pump seals, and reagent preparation. Most of the water is lost through evaporation. Seawater scrubbing is the exception, since the flue gas is cooled below its dew point and the condensation of moisture within the gas actually results in a net water gain in the scrubber. Since the seawater process is operated in a once-through mode, all scrubbing liquor and moisture condensed from the flue gas are discharged as blowdown. The spray dryer process loses less water to evaporation than other FGD processes since the flue gas is not saturated.

Water occluded in the solid waste represents the second largest loss from most processes, although blowdown of scrubbing liquor can also be significant if the process is not designed for zero-discharge. Natural oxidation processes typically lose sufficient scrubbing liquor due to occlusion in the solid waste that operation with zero liquid waste discharge is attainable. However, the improved dewatering characteristics of the gypsum sludge produced by forced oxidation processes means that less water is lost due to occlusion in the solid waste. This can make zero-discharge operation difficult to achieve

TABLE S-1 SUMMARY OF COAL TYPES AND FGD SYSTEMS EVALUATED

Coal Type*	Typical of	Coal Proximate Analysis				FGD Processes	SO_2 Removal Required*** (%)
		S (%)	Ash (%)	Moisture (%)	AHV** (MJ/kg)		
1	Eastern High Sulphur Bituminous (New Brunswick)	7.0	20	5	25	- Limestone - Wellman-Lord	90
2	Eastern Medium Sulphur Bituminous (Nova Scotia)	4.0	12	7	28	- Limestone - Seawater Scrubbing	90
3	Western Lignite (Saskatchewan)	0.5	13	25	15	- Limestone - Spray Dryer - Alkaline Fly Ash	60
4	Ontario	2.0	8.5	7	29.6	- Limestone - Lime - Dual Alkali - Wellman-Lord	80

* Western subbituminous coal not evaluated.

** Average heat value.

*** SO_2 removal required to meet National Guidelines for Atmospheric Emissions from New Thermal Power Generation Facilities, assuming 92 percent of S in fuel evolves as SO_2.

TABLE S-2 WATER LOSSES FOR FGD PROCESSES[a]

FGD Process	Coal Type[b]	Water Losses (L/min) Evaporation in Scrubber	Occlusion in Solid Waste[c]	Entrainment in Product Stream	Blowdown	Total
1) Limestone						
Without Forced Oxidation	1	1 856	697	-	-	2 553
	2	1 518	330	-	-	1 848
	3	1 968	58	-	-	2 026
	4	1 543	155	-	-	1 698
With Forced Oxidation	1	1 856	464	-	-	2 320
	2	1 518	220	-	-	1 738
	3	1 968	39	-	-	2 007
	4	1 543	103	-	-	1 646
2) Spray Dryer	3	1 782	110	-	-	1 892
3) Alkaline Fly Ash	3	1 968	328	-	-	2 296
4) Seawater	2	(840)[d]	-	-	838 585	837 745
5) Wellman-Lord	1	1 856	54[e]	38	92	2 040
	4	1 543	12[e]	18	33	1 606
6) Lime						
Without Forced Oxidation	4	1 543	141	-	-	1 684
With Forced Oxidation	4	1 543	97	-	-	1 640
7) Dual Alkali	4	1 543	141	-	-	1 684

[a]Water losses are for a 500-MW(e) station operating at 37% energy conversion efficiency.

[b]Coal Type

		% S	AHV (MJ/kg)
1	High S Bituminous	7.0	25
2	Med. S Bituminous	4.0	28
3	Western Lignite	0.5	15
4	Ontario	2.0	29.6

[c]Includes water of hydration.

[d]Moisture condensed from flue gas.

[e]Water evaporated in purge solids dryer.

by reducing the quantity of fresh water that can be introduced for mist eliminator washing, etc., and by increasing the dissolved solids concentration in the scrubbing liquor to the point that materials corrosion becomes severe. The result may be a need to operate forced oxidation systems with a blowdown.

Systems equipped with a prescrubber loop for control of suspended solids and chlorides, such as the Wellman-Lord process, also generally operate with a blowdown.

Solid Waste

Table S-3 summarizes the quantities of ash and FGD solids that would be generated by a hypothetical 500-MW station for each coal type/FGD system case considered. This table illustrates that, whereas forced oxidation produces more dry solids (by weight) than natural oxidation processes, the total volume of wet FGD sludge ultimately requiring disposal is less due to the better dewatering properties of the former.

For a given coal type, the spray drying process produces a lower volume of solid waste than any of the wet lime or limestone based processes. This is due to the lower water content of the solid waste from the spray dryer.

Seawater scrubbing does not produce a solid waste as such, because the sulphur reaction products are discharged as dissolved solids in the scrubber blowdown stream. Residual fly ash scrubbed in the absorber and inert material are released as suspended solids in the blowdown.

The regenerable Wellman-Lord process produces only a fraction of the quantity of waste solids generated by the throwaway processes, since most of the sulphur dioxide scrubbed from the flue gas is converted to sulphur by-products.

Waste Characteristics

The relevant parameters of Environment Canada's proposed criteria for discharge of wastewaters from thermal power generating stations are summarized below.

pH	6.5 - 9.5
Chromium	0.5 mg/L
Copper	0.5 mg/L
Iron	1.0 mg/L
Nickel	0.5 mg/L
Zinc	0.5 mg/L
Suspended Solids	25 mg/L

TABLE S-3 SOLID WASTE PRODUCTION RATES AND DISPOSAL VOLUMES[a]

FGD Process	Coal Type[b]	Solid Waste Production (kg/min) FGD Solids (Dry)	FGD Sludge[c] (As Disposed)	Bottom Ash[d] (Dry)	Fly Ash[d] (Dry)	20-Year Landfill Volume (10^6 m^3) FGD Sludge[e]	Bottom Ash[f]	Fly Ash[f]	Total
Limestone									
Without Forced Oxidation	1	931	1551	162	486	6.89	0.77	2.31	9.97
	2	441	735	81	243	3.07	0.38	1.14	4.59
	3	78	130	175.3	526	0.58	0.83	2.49	3.90
	4	207	345	58.3	175	1.53	0.28	0.84	2.65
With Forced Oxidation	1	1107	1384	162	486	5.04	0.77	2.31	8.12
	2	524	656	81	243	2.39	0.38	1.14	3.91
	3	93	116	175.3	526	0.42	0.83	2.49	3.74
	4	247	309	58.3	175	1.12	0.28	0.84	2.24
Alkaline Fly Ash	3	577[g]	888	175.3	-	3.77	0.83	-	4.60
Spray Dryer	3	578[g]	680	175.3	-	2.35	0.83	-	3.18
Seawater	2	-	-	81	243	-	0.38	1.14	1.52
Wellman-Lord	1	81	81	162	486	0.23	0.77	2.31	3.31
	4	18	18	58.3	175	0.05	0.28	0.84	1.17
Lime									
Without Forced Oxidation	4	187	312	58.3	175	1.38	0.28	0.84	2.50
With Forced Oxidation	4	222	278	58.3	175	1.01	0.28	0.84	2.13
Dual-Alkali	4	187	312	58.3	175	1.38	0.28	0.84	2.50

[a]For hypothetical 500-MW station operating at 37 percent energy conversion efficiency, 65 percent load factor.

[b]Coal Type

		% S	AHV (MJ/kg)
1	High S Bituminous	7.0	25
2	Med. S Bituminous	4.0	28
3	Western Lignite	0.5	15
4	Ontario	2.0	29.6

[c]Limestone and lime sludge without forced oxidation is assumed to contain 60 percent solids. Limestone and lime sludge with forced oxidation is assumed to contain 80 percent solids.

[d]Bottom ash/fly ash split = 25 percent/75 percent.

[e]Specific gravity of 2.40 g/cm^3 was assumed for all dry FGD solids.

[f]Dry disposal of bottom ash and fly ash assumed. Dry bulk density was assumed to be 1.44 g/cm^3.

[g]FGD solids include fly ash.

Laboratory analyses performed on process wastewaters, potential wastewaters (e.g., disposal pond return water), and runoff from solid waste disposal sites indicated that the characteristics of these streams generally fell within the proposed discharge criteria with regard to chromium, copper, iron, nickel, and zinc concentrations, although pH and suspended solids criteria were frequently exceeded. Treatability studies indicated that lime coagulation produced an effluent that met the proposed discharge criteria for one sample in which the criteria for iron and nickel concentrations were exceeded.

Although these are the only current federal guidelines which would apply to FGD wastewater discharges, other parameters such as chemical oxygen demand (COD), total dissolved solids (TDS), and other heavy metals may be of environmental concern, and may require treatment to meet other regulatory requirements.

The physical properties of FGD sludge are affected by the extent of oxidation, amount of fly ash present and water content. Highly oxidized calcium sludges dewater readily to a structurally stable material (gypsum) with physical properties superior to the thixotropic characteristics of predominantly sulphite sludges. Reducing the water content of sulphite sludges and mixing them with alkaline fly ash and commercial additives can significantly improve their structural properties.

The chemical characteristics of FGD sludges are more strongly influenced by the scrubbing reagent used and the coal and fly ash compositions. Leachates from sludges of sodium-based FGD processes exhibit higher TDS concentrations than those of calcium-based systems, due to the relatively higher solubility of sodium salts. FGD sludges examined in this study from western U.S. stations burning subbituminous and lignite coals generally contained higher concentrations of most metals than those from eastern bituminous coals. This may be due to differences in the amount and composition of fly ash contained in the sludge.

FGD Process Design Options

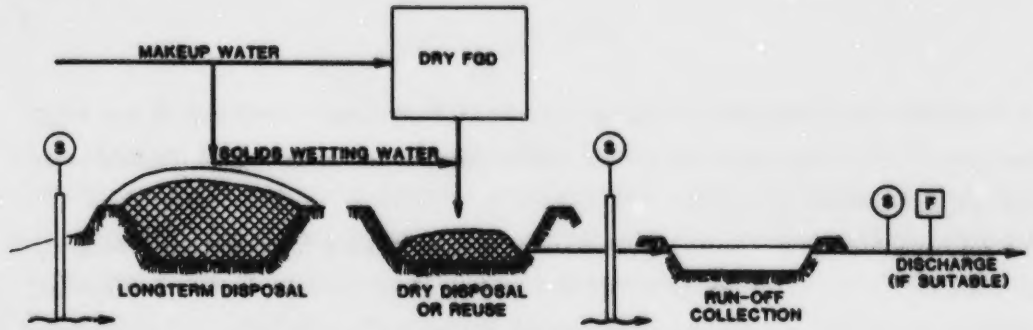
Several opportunities exist in the design of an FGD system to incorporate features that will improve process reliability and minimize its environmental impact. Some of those identified in this study are summarized below.

1. Dry Scrubbing - Spray dryer FGD systems offer simplicity of operation, the elimination of any process wastewater, and the production of a dry, manageable solid waste with good disposal properties and reuse potential. Consideration should be given to selection of this process, particularly for low sulphur coal applications.

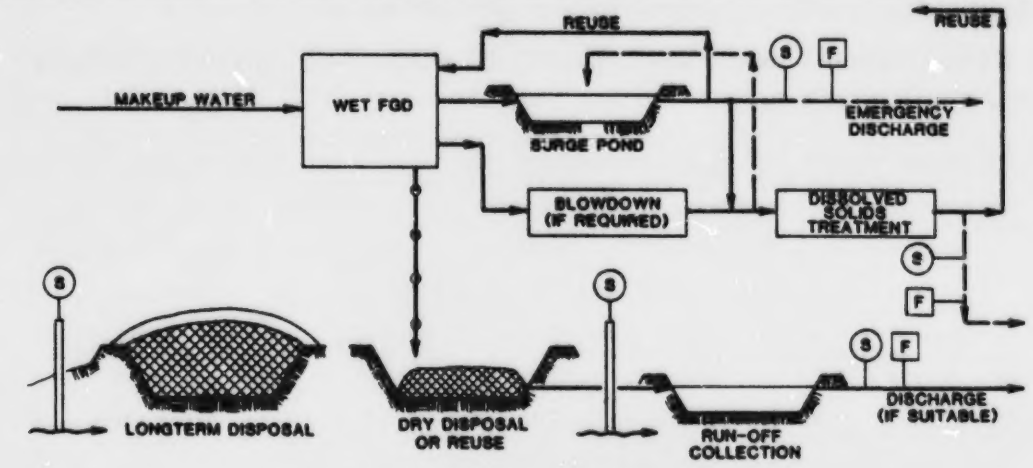
2. Cascade Water Use - The use of station wastewater streams as makeup to the FGD system should be considered to minimize water use and the resulting entrainment of aquatic organisms.
3. Solid Waste Conditioning - Solid waste should be conditioned to produce a relatively stable product in terms of its physical and chemical properties, prior to disposal. This would normally entail either the use of forced oxidation to produce calcium sulphate or the fixation or stabilization of predominantly sulphite sludges and dry scrubbing wastes.
4. Solid Waste Disposal - Solid waste disposal areas should be designed in accordance with the following principles:

   i) Provision should be made to reclaim the site upon abandonment. This necessitates the placement of a physically and chemically stable material in the disposal area throughout its lifetime. In the case of ponds, provision should be made to dewater the contents after interim storage of the sludge.
   ii) Disposal areas should be designed to minimize contact between waste solids and water. Low permeability containment should be provided to prevent filtration of groundwater and to restrict seepage of leachate. Sites should be reclaimed in a modular fashion to minimize the amount of material exposed to the effects of precipitation and thereby limit runoff.
   iii) The disposal area should provide containment for all process flows and infiltration (precipitation and runoff) that it receives.

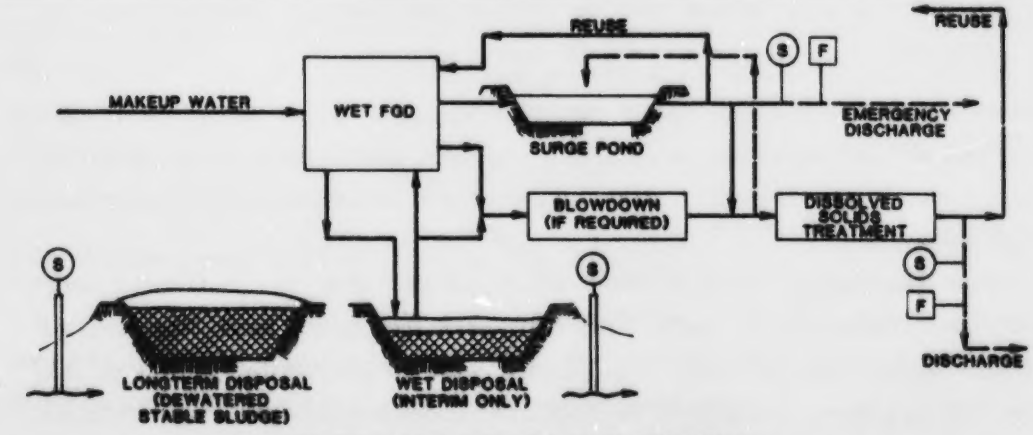
   As these objectives are more readily met with dry disposal practices, landfilling is environmentally preferred to wet disposal methods, particularly in areas of net precipitation. Figure S-1 illustrates some alternate waste management practices for FGD solid waste.
5. Wastewater Discharge - FGD systems should be provided with sufficient surge capacity to accommodate upsets, equipment drainings, and runoff from process areas and waste disposal sites. Compliance with environmental guidelines may require that no process wastewater be discharged under normal operating conditions. While this has been accomplished with spray dryer and natural oxidation wet scrubbing processes, further work is required to investigate the technical and cost implications of closed-loop operation with forced oxidation systems. Flow monitoring and sampling facilities should be provided for emergency wastewater



A) DRY FGD AND DRY DISPOSAL



B) WET FGD AND DRY DISPOSAL



C) WET FGD AND INTERIM WET DISPOSAL

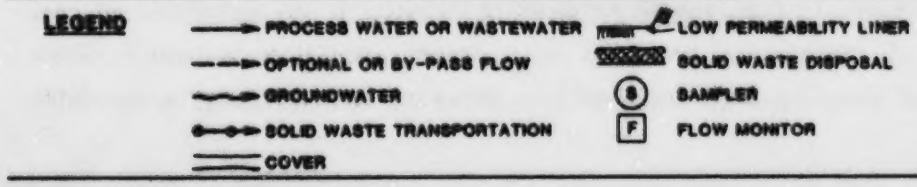


FIGURE 5-1 ALTERNATIVE FGD WASTE MANAGEMENT DESIGNS

discharges. Groundwater in the vicinity of the solid waste disposal site should be monitored on a routine basis. Figure S-1 illustrates some alternative water management designs for closed loop operation of FGD systems.

6. Waste Utilization - Prospects for FGD waste utilization should be considered. The use of by-product gypsum from forced oxidation limestone processes appears to be the most economically feasible option at this time, although the technical and cost implications require further investigaton.
7. Reliability - In the past decade, significant advancements have been made in improving the historically low reliability of FGD systems. This has been accomplished largely through increasing equipment redundancy and by closer attention to the selection of materials of construction. Since these improvements in reliability have generally resulted in lower operation and maintenance costs at the expense of a higher first-cost for the system, a detailed analysis of the sparing philosophy and materials selection is warranted at the design phase.

Costs

Within the accuracy of the cost data compiled in this generic study, significant differences cannot be detected between the wet scrubbing, throwaway processes considered. However, Wellman-Lord, the only regenerable process examined, appears to have significantly higher capital and operating costs, even when a full cost credit for by-product sulphur is included. Forced oxidation with landfill of waste solids appears to offer some reduction in capital cost but may have marginally higher annual revenue requirements than natural oxidation with ponding of untreated sludge. The use of organic acid enhanced limestone scrubbing may offer additional cost savings.

1 INTRODUCTION

1.1 Background

In April 1981, the Federal Minister of the Environment issued national guidelines for the control of atmospheric emissions from new fossil fuel-fired power generating stations. These guidelines include limits for the emission of sulphur dioxide (SO_2) and other pollutants (1). Provincial regulatory agencies are also developing sulphur dioxide emission standards, guidelines and regulations. These guidelines are being developed, in part, in response to the considerable scientific and public concern which has been expressed in recent years regarding the contribution that sulphur dioxide emissions from Canadian coal-fired power stations have made to the acid rain problems experienced in North America.

One control option which has been commercially demonstrated in the United States, Europe, and Japan is flue gas desulphurization (FGD). As Canadian utility companies and regulatory agencies formulate strategies to control sulphur dioxide emissions, it is anticipated that the three principal criteria used to critically assess the available FGD options will be:

a) environmental implications,
b) system reliability, and
c) relative cost.

For different types of FGD systems, and different design options for these systems, the volume of makeup water required and the volume and characteristics of wastewater and solid wastes produced can vary significantly. These variations in turn affect the level of environmental concern and the cost of environmental controls. Environment Canada views the early stage of FGD system implementation in Canada as an opportunity to identify those systems and design options that will minimize water-related and solid waste disposal concerns in Canadian situations.

1.2 Objectives

The principal objective of this study was to critically evaluate the commercial demonstrated FGD alternatives currently available, with regard to their environmental implications due to wastes produced, their costs and reliability. This report encompasses the findings of a two-phase study. Phase One focused on possible FGD developments in Ontario. Phase Two consisted of a similar evaluation for the rest of Canada.

The specific objectives of the study were as follows:

1) to evaluate the most feasible FGD systems, the design options for those systems, and the alternative solid waste and wastewater management strategies available, for different Canadian regional areas and coal types;
2) to characterize chemically, physically and volumetrically the wastewaters and solid wastes from appropriate FGD systems and designs;
3) to investigate environmentally responsive sludge treatment and containment techniques;
4) to conduct a preliminary market and technical survey of waste utilization options for FGD waste products;
5) to identify practicable wastewater and solid waste disposal control methods, options and guidelines; and
6) to estimate and compare the costs of solid waste and wastewater management for the various options and the effects of these costs on total system costs.

1.3 Methodology

Utility companies in the United States have developed extensive design and operating experience with FGD systems applied to some of their coal-fired electrical generating stations. A literature review was conducted to identify operating FGD systems and design alternatives in the U.S.A. that were considered applicable to coal types and other factors representative of Canadian situations. Power stations in the United States using such systems were then visited.

The objectives of the site visits were two-fold. First, engineering and operating data were compiled for each facility and discussions were held with station personnel to identify the problems and successes they had encountered in their experience with the FGD system. Second, samples of some sludges and wastewaters were collected for waste characterization. Sludge leaching experiments and laboratory analyses were conducted to identify the characteristics of the sludges and wastewaters which could be generated by the processes considered.

Using the above information, the wastewater and solid waste management alternatives for each FGD system considered were evaluated. Mass balances were developed for a model 500-MW station to compare the makeup water and reagent requirements and the wastewater and solid waste volumes generated by each process. Costs of each FGD process and associated waste management alternatives were examined and the reliability of FGD systems was investigated.

The study was concluded with the identification of environmentally preferred FGD system types, design options, and waste management practices applicable to Canadian coal-fired generating stations.

2 COAL-FUELLED GENERATING STATIONS IN CANADA

2.1 Introduction

Coal-fuelled generating stations in Canada have a wide range of generating capacities and are located in geographic areas that experience diverse climatic conditions. The generating stations utilize a variety of coal ranks which may differ markedly in their physical-chemical properties.

2.2 Locations, Capacities and Coal Use at Existing Coal-Fuelled Generating Stations

Data for Ontario stations, which were considered in the first phase of this study, are presented in Figure 1 and Table 1 (2). The following discussion deals primarily with the remainder of the coal-fired stations in Canada.

As shown in Figure 2, the Western Canadian generating stations are located in Alberta, Saskatchewan and Manitoba; and the Maritime stations are in New Brunswick and Nova Scotia. The eastern stations are situated close to communities and, for the most part, are on the ocean. These stations have relatively smaller generating capacities, and are generally older, oil and coal-fired units. The newer, larger, western stations are usually located adjacent to a mine mouth, rather than at an electrical load centre.

A summary of the locations, generating capacities, annual fuel consumption, and types and origins of coals used at existing Maritime and Western Canadian stations is presented in Table 2(3). Approximately 30 percent of the total generating capacity is located in Alberta (the largest coal producing province) and Saskatchewan, about 10 percent in the Maritimes and Manitoba, and about 60 percent in Ontario. Coal consumption is highest in Ontario generating stations, followed by those in Alberta, Saskatchewan, Nova Scotia, New Brunswick, and Manitoba (3).

Bituminous ranked coal represents about 53 percent of the annual amount of coal burned in Canada, and is utilized predominantly in the Maritimes and Ontario. Subbituminous coal represents 31 percent of the coal burned in Canada, primarily in Alberta. Lignite represents about 14 percent, which is used at the Saskatchewan and Manitoba stations. At the Milner Generating Station in Alberta, rejects and dewatered tailings from a metallurgical grade processing plant are used as the primary fuel.

Generally, the rate of growth of Canadian electrical generating facilities has decreased in recent years. In the future, additional coal-fuelled stations are expected to be installed in Alberta, Saskatchewan and Nova Scotia. In the longer term, new coal-

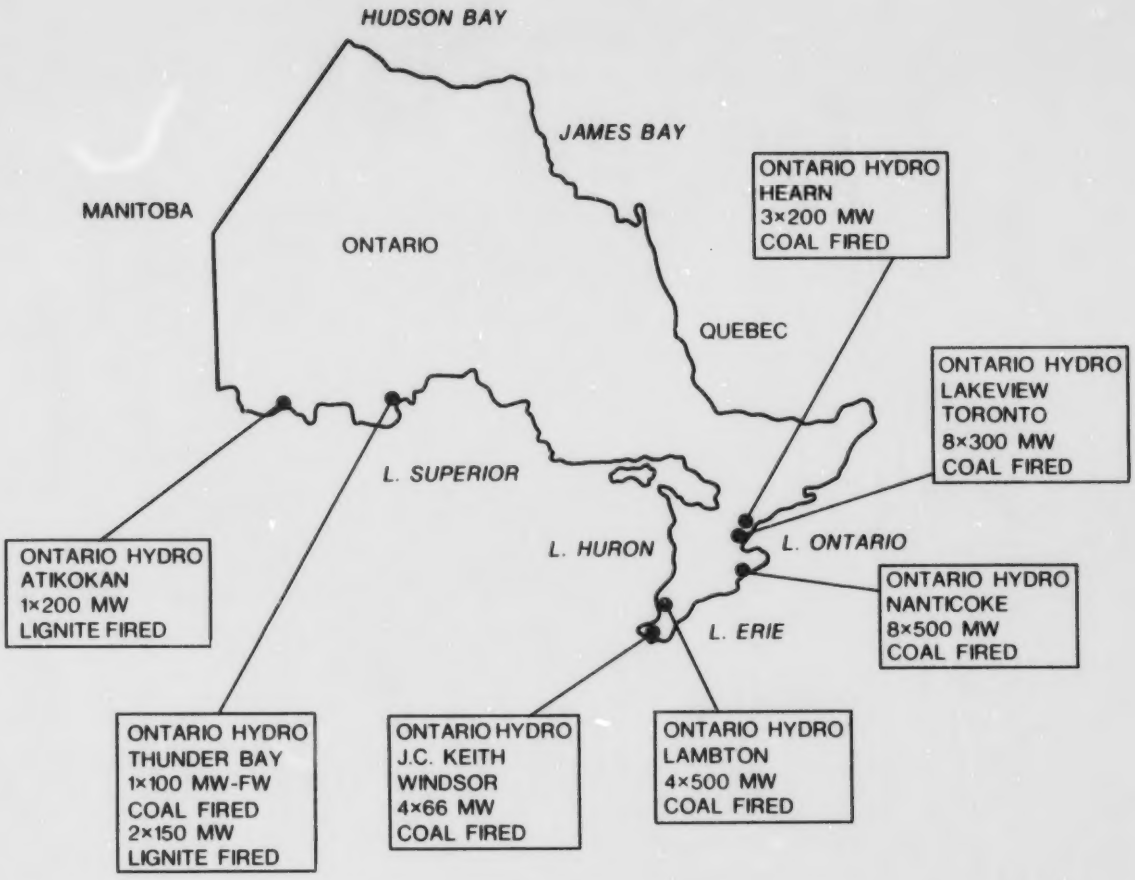


FIGURE 1 LOCATION AND GENERATING CAPACITIES OF ONTARIO HYDRO'S COAL FIRED STEAM ELECTRIC STATIONS (2)

fuelled stations may also be installed in New Brunswick, Ontario, British Columbia and Manitoba.

2.3 Properties of Coal Used

Analyses of coals used in Ontario and in Western Canadian and Maritime generating stations are presented in Tables 3 and 4, respectively. A combination of American bituminous and western Canadian bituminous and lignite is used in Ontario (2). In general, the coals used in the west have a substantially higher moisture content and ash alkalinity than the Maritime coals. These coals also have a somewhat lower heating value, corresponding to a lower percentage of fixed carbon, volatile matter and ash. Furthermore, the western coals have a significantly lower sulphur content (about 10 percent of the sulphur content of the Maritime coals).

TABLE 1 GENERATING CAPACITIES AND COAL USE AT ONTARIO HYDRO COAL FIRED STEAM ELECTRIC STATIONS (2)

		Annual - 1981					Projected 1990		
Generating Stations	Location	Rated Capacity (MW)	Average Capacity Factor*	Coal Consumption (10^6 tonnes)	Coal Type	%	Average Capacity Factor*	Coal Consumption (10^6 tonnes)	Coal Type
Hearn	Toronto	600	9	0.191	U.S. Medium Sulphur Bituminous	100	1	0.020	U.S. Medium Sulphur Bituminous
Lakeview	Mississauga	2400	25	1.694	U.S. Medium Sulphur Bituminous	100	10	0.680	U.S. Medium Sulphur Bituminous
Nanticoke	Nanticoke	4000	47	3.409	U.S. Regular Sulphur Bituminous	60	25	1.810	U.S. Regular to Medium Sulphur Bituminous
				2.364	Western Canadian Bituminous	40		1.26	Western Canadian Bituminous
Keith	Windsor	264	16	0.093	U.S. Medium Sulphur Bituminous	100	1	0.006	U.S. Medium Sulphur Bituminous
Lambton	Courtright	2000	57	2.926	U.S. Regular Sulphur Bituminous	89		2.260	U.S. Medium Sulphur Bituminous
				0.356	U.S. Low Sulphur Bituminous	11	44**	0.275	U.S. Low Sulphur Bituminous
Thunder Bay	Thunder Bay	400	49	0.144	U.S. Regular Sulphur Bituminous	34	Unit 1 - 1	0.0035	U.S. Medium Sulphur Bituminous
				0.280	Western Canadian Lignite	66	Unit 2&3-64	0.720	Western Canadian Lignite
Atikokan	Atikokan	200	0	0	Western Canadian Lignite	100	65	0.810	Western Canadian Lignite

* Average Capacity Factor (ACF) = $\frac{\text{Operating Hours x Average MW}}{\text{8760 x Maximum Rated MW}} \times 100$

** FGD retrofitted units 3 and 4 would have an ACF of about 60%. Unscrubbed units 1 and 2 would have an ACF of about 30%.

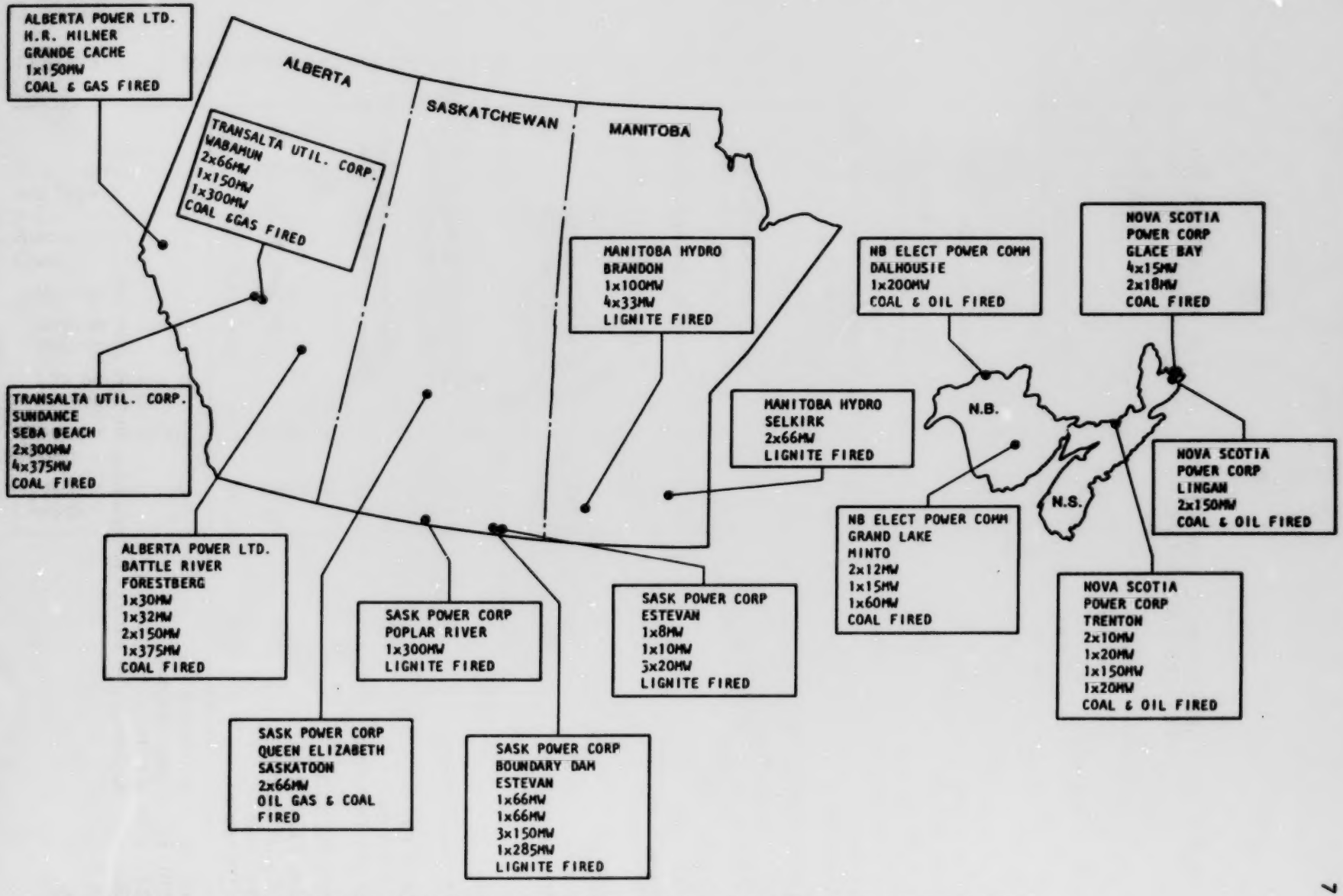


FIGURE 2 LOCATIONS AND GENERATING CAPACITIES OF COAL-FUELLED GENERATING STATIONS IN THE MARITIMES AND WESTERN CANADA (4)

TABLE 2 GENERATING CAPACITY AND COAL USE AT EASTERN AND WESTERN CANADIAN COAL FIRED GENERATING STATIONS (3)

Generating Station (Location)	Rated Capacity (MW)	Coal Consumption (10^6 tonnes)	Coal Origin (Type)
Alberta			
Battle River (Forestburg)	737	2.5	Forestburg, Alberta (SB)
Milner (Grande Cache)	150	0.34	Grande Cache, Alberta (R)
Sundance (80 km West of Edmonton)	2100	6.7	Wabamum, Highvale Mine, Alberta (SB)
Wabamum (Wabamum)	582	1.5	Wabamum, Whitewood Mine, Alberta (SB)
Saskatchewan			
Queen Elizabeth (Saskatoon)	232	0.36	Sheerness, Alberta (SB)
Estevan (Estevan)	67	0.44	Estevan, Saskatchewan (L)
Boundary Dam (Estevan)	882	4.2	Estevan, Saskatchewan (L)
Poplar River (Coronach)	300	1.9	Willowbunch, Saskatchewan (L)
Manitoba			
Selkirk (Selkirk)	156	0.023	Estevan, Saskatchewan (L)
Brandon (Brandon)	237	0.023	Estevan, Great West Coal, Sask. (L)
Nova Scotia			
Trenton (Trenton)	210	0.339	75% Cape Breton, Nova Scotia (B); 25% Other Sources
Glace Bay (Glace Bay)	112	0.39	Cape Breton, Nova Scotia (B)
Lingan (Dalhousie)	300	0.518	Cape Breton, Lingan, N.S. (B)
New Brunswick			
Grand Lake (New Castle Creek)	104	0.176	Minto, New Brunswick (B)
Dalhousie (Dalhousie)	212	0.272	Minto, N.B. (B)

(B) : Bituminous
(SB) : Subbituminous
(L) : Lignite
(R) : Rejects and dewatered tailings from metallurgical coal.

TABLE 3 ANALYSIS OF COAL USED BY ONTARIO HYDRO IN 1981 (2)

Coal Source and Type	Quantity Used (tonnes)	As Received Analysis					
		Moisture (%)	Ash (%)	Volatile (%)	Fixed Carbon (%)	Sulphur (%)	Heating Value (MJ/kg)
Average U.S. Coal	8.82×10^6	5.71	8.88	35.37	50.04	2.41	30.0
Regular	6.48×10^6	5.57	9.91	35.38	49.12	2.82	29.6
Medium Sulphur Bituminous	1.98×10^6	5.92	7.86	34.55	51.67	1.72	30.5
Low Sulphur Bituminous	0.356×10^6	7.86	6.89	31.89	53.36	0.81	30.3
Western Canadian Bituminous	2.364×10^6	9.92	10.81	30.33	48.94	0.25	25.2
Western Canadian Lignite	0.280×10^6	32.3	12.88	24.5	30.32	0.46	15.5

TABLE 4 PROXIMATE ANALYSES OF COAL USED BY GENERATING STATIONS IN WESTERN CANADA AND THE MARITIMES (3,4,5,6,7)

Generating Station	Moisture (%)	Ash (%)	Volatile (%)	Fixed Carbon (%)	Sulphur (%)	Heating Value (MJ/kg)	Ash Alkalinity (CaO Content) (%)
Alberta							
Battle River	27.0	11.0	27.8	34.2	0.4	20.8	16.4
Milner	7.2	40.2	13.4	39.2	0.4	21.5	8.6
Sundance	20.4	13.6	29.5	36.5	0.21	19.9	13.1
Wabamum	21.4	13.5	29.0	36.1	0.23	19.3	11.9
Saskatchewan							
Queen Elizabeth	27.6	9.0	30.3	31.1	0.5	19.3	12.2
Estevan	35.2	10.7	25	31	0.5	16.3	14.6
Boundary Dam	37.5	9.1	24.2	28.2	0.4	15.6	11.8
Poplar River	38	13.1	24.5	23.9	0.6	12.8	NA*
Manitoba							
Selkirk	33.5	9.0	25.4	32.2	0.5	16.3	14.8
Brandon	33.4	7.9	26.3	32.4	0.4	16.0	14.8
Nova Scotia							
Trenton	6.9	19.7	30.2	43.2	4.7	24.0	1.0
Glace Bay	6.0	10.1	32.5	56.8	1.6	32.1	1.8
Lingan	7	12	33	55	4.0	29.9	1.8
New Brunswick							
Grand Lake	2	13.2	32	48	4-8	27.5	3.1
Dalhousie	9	24	27	40	4-8	23.3	6.0

* Not Available.

2.4 Climatic Factors

Regional climatic factors are a consideration in the design of FGD waste management systems. General climatic conditions for Ontario are presented in Table 5 (2). Climatic conditions for the three western provinces, Alberta, Saskatchewan, and Manitoba and two Maritime provinces, New Brunswick and Nova Scotia, are presented in Table 6. Influenced by topography and relation to the sea, the Maritime sites generally receive over twice as much precipitation as the inland sites in the western provinces. The Atlantic coast receives most of its precipitation in the winter, unlike the interior plains region where most precipitation occurs in the summer (8). In both regions rainfall exceeds the snow water equivalent. The stations located in the Maritimes experience milder winter temperatures than those in the west but similar summer temperatures.

The mean lake evaporation is somewhat higher in the western provinces and, except for the site of the Milner Station, exceeds annual precipitation. However, the mean annual runoff, influenced by factors such as precipitation, topography and latitude, is vastly greater at the eastern sites, with the highest occurring in Nova Scotia and the lowest in Saskatchewan, Manitoba, and parts of Alberta.

TABLE 5 CLIMATIC DATA FOR ONTARIO HYDRO STATIONS (2)

Generating Stations	Mean Annual Precipitation[1] (mm)	Mean Annual Snow Fall[2] (cm)	Mean January Daily Temperature (°C)	Mean July Daily Temperature (°C)	Mean Lake Evaporation[3] (mm)	Derived Annual Runoff[4] (mm)
Hearn	720	100	-5	21	800	200
Lakeview	750	105	-5	22	800	300
Nanticoke	945	150	-5.6	20.6	800	300
Keith	800	80	-4.2	22.8	800	200
Lambton	920	210	-9.8	21	800	300
Thunder Bay	740	210	-15.4	17.6	500	300
Atikokan	785	240	-18.4	17.2	600	300

1) Mean Annual Precipitation is the accumulation of rainfall and the water equivalent of snowfall and intermediate forms of precipitation (e.g., hail).
2) The water equivalent of Mean Annual Snowfall (approximately 0.10 g/cm^3) is included in mean precipitation.
3) Mean Lake Evaporation is empirically determined from evaporation pan data supplemented by calculations based on climatological data.
4) Derived Annual Runoff is an estimate of areal runoff calculated by subtracting estimated annual evapotranspiration from annual precipitation. The relationship precipitation - evaporation = runoff cannot be rigorously applied to the above data.

TABLE 6 CLIMATIC DATA FOR SITES OF COAL-FUELLED GENERATING STATIONS IN THE MARITIMES AND WESTERN CANADA (8)

Generating Stations	Mean Annual Precipitation[1] (mm)	Mean Annual Snowfall[2] (cm)	Mean January Temperature (°C)	Mean July Daily Temperature (°C)	Mean Lake Evaporation[3] (mm)	Derived Annual Runoff[4] (mm)
Alberta						
Battle River	370	127	-15.8	17.0	705	23
Milner	590	280	-15	13.9	400	500
Sundance	520	150	-15	15.2	565	125
Wabamum	510	145	-15.6	15.3	570	115
Saskatchewan						
Queen Elizabeth	360	103	-18.7	18.6	760	10
Estevan	520	107	-16.7	18.9	870	23
Boundary Dam	515	104	-16.4	19.4	880	24
Poplar River	315	80	-14.3	18.7	900	25
Manitoba						
Selkirk	530	136	-18.8	19.7	650	47
Brandon	490	128	-17.8	18.8	770	50
Nova Scotia						
Trenton	1130	220	-7.5	17.5	570	800
Glace Bay	1400	240	-3.4	17.5	540	1100
Lingan	1400	220	-3.1	17.5	545	1050
New Brunswick						
Grand Lake	1160	240	-8.1	17.5	565	675
Dalhousie	100	320	-11.3	17.5	505	570

1) Mean Annual Precipitation is the accumulation of rainfall and the water equivalent of snowfall and intermediate forms of precipitation (e.g., hail).
2) The water equivalent of Mean Annual Snowfall (approximately 0.10 g/cm^3) is included in mean precipitation.
3) Mean Lake Evaporation is empirically determined from evaporation pan data supplemented by calculations based on climatological data.
4) Derived Annual Runoff is an estimate of areal runoff calculated by subtracting estimated annual evapotranspiration from annual precipitation. The relationship precipitation - evaporation = runoff cannot be rigorously applied to the above data.

3 FGD PROCESSES

3.1 Introduction

Flue gas desulphurization (FGD) is the generic term applied to a number of processes that remove sulphur dioxide (SO_2) from combustion gases. These processes all use a reagent (sorbent) to absorb and react with the SO_2 to form compounds that are either treated to regenerate the sorbent and produce marketable by-products (e.g., elemental sulphur, sulphuric acid, or concentrated SO_2), or are disposed of or reused in their existing form without sorbent regeneration. Those processes where the sorbent is regenerated and reused are referred to as regenerable processes; other processes are termed non-regenerable or throwaway processes.

FGD processes may be further classified as either wet or dry scrubbing systems. With wet systems, the flue gas is scrubbed with a circulating alkaline solution or slurry at sufficiently high liquid-to-gas (L/G) ratios to quench the gas to its adiabatic saturation temperature. Sulphur dioxide is absorbed into the scrubbing liquor where it reacts with the sorbent to produce sulphur compounds which must ultimately be separated from the liquid phase. Wet systems are characterized by a number of liquid handling systems for reagent preparation, absorption, and regeneration or solids disposal. Dry scrubbing systems, on the other hand, operate above the flue gas saturation temperature and no liquid stream exits the scrubber. They use an alkaline-based sorbent which may either be dry-injected directly into the gas stream or sprayed as a solution or slurry at sufficiently low L/G ratios to ensure that all water is evaporated, leaving only dry reaction products which are then removed in particle collectors.

Table 7 summarizes the breakdown of operational, under construction, and planned FGD systems in the U.S. utility industry as of January 1982 (9). The vast majority of FGD systems used to date for scrubbing utility boilers in North America have been wet, non-regenerable processes. In fact, lime and limestone processes (including alkaline fly ash) represent over 85 percent of the total operational FGD capacity in the U.S. However, a recent trend in the ordering of new systems has been towards the spray drying process, which represents approximately 14 percent of system capacity currently under construction, compared to only 1.7 percent of current operational capacity. Regenerable processes represent less than 5 percent of the FGD operational capacity in the U.S. and do not appear to be increasing.

This section reviews the process fundamentals of six FGD processes considered to have the greatest potential for application in Canada within the next 10 years, because

TABLE 7 DISTRIBUTION OF U.S. UTILITY FGD SYSTEMS BY PROCESS AS OF JANUARY 1982 (9)

Process	FGD Capacity							
	Operational		Under Construction		Planned		Total	
	MW	% of Total	MW	% of Total	MW	% of Total	MW	Total
Non-Regenerable:								
Limestone*	15 578	47.7	8 179	49.1	14 085	61.7	37 842	52.4
Lime**	12 499	38.2	4 270	25.6	4 940	21.6	21.709	30.1
Lime/Limestone	20	0.1	0	0	475	2.1	495	0.7
Lime/Sodium Carbonate	0	0	100	0.6	0	0	100	0.1
Dual Alkali (Lime and Limestone)	1 181	3.6	421	2.5	421	1.8	2 023	2.8
Sodium Carbonate	1 255	3.8	0	0	1 900	8.3	3 155	4.4
Spray Drying	550	1.7	2 338	14.0	1 019	4.5	3 907	5.4
Regenerable:								
Wellman-Lord	1 540	4.7	534	3.2	0	0	2 074	2.9
Citrate	60	0.2	0	0	0	0	60	0.1
Aqueous Carbonate	0	0	100	0.6	0	0	100	0.1
Magnesium Oxide	0	0	724	4.4	0	0	724	1.0
Total	32 683	100.0	16 666	100.0	22 840	100.0	72 189	100.0

* Includes Limestone Alkaline Fly Ash Systems.

** Includes Lime Alkaline Fly Ash Systems.

of their level of commercial experience, and applicability with Canadian coals. These six processes are:

- Limestone
- Lime
- Dual Alkali
- Alkaline Fly Ash
- Spray Drying
- Wellman-Lord

Other processes briefly addressed because of their unique design features are:

- Sodium Carbonate
- Seawater Scrubbing
- Davy-Saarberg Hoelter Process
- Dry Injection
- Magnesium Oxide
- Aqueous Carbonate
- Citrate

3.2 Limestone Scrubbing Process

A simplified schematic of the commonly used limestone scrubbing process is presented in Figure 3. Following particulate removal, flue gas typically enters near the bottom of an absorber tower and flows upward, countercurrent to the flow of scrubbing liquor. Although packed, venturi, spray, tray and turbulent contact absorbers have all been used, conventional spray towers have gained greatest popularity in this application due to their relatively low pressure drop and absence of complicated internals which tend to aggravate plugging and scaling problems (10).

The scrubbing liquor consists of a slurry of finely ground (200 to 300 mesh) limestone ($CaCO_3$) which absorbs SO_2 and reacts to produce calcium sulphite ($CaSO_3$) and calcium sulphate ($CaSO_4$) solids according to the following simplified reaction chemistry (11):

SO_2 Absorption/Dissolution:	$SO_2 + H_2O \rightarrow H_2SO_3$
Neutralization/Precipitation:	$H_2SO_3 + CaCO_3 \rightarrow CaSO_3 + CO_2 + H_2O$
Oxidation:	$2CaSO_3 + O_2 \rightarrow 2CaSO_4$

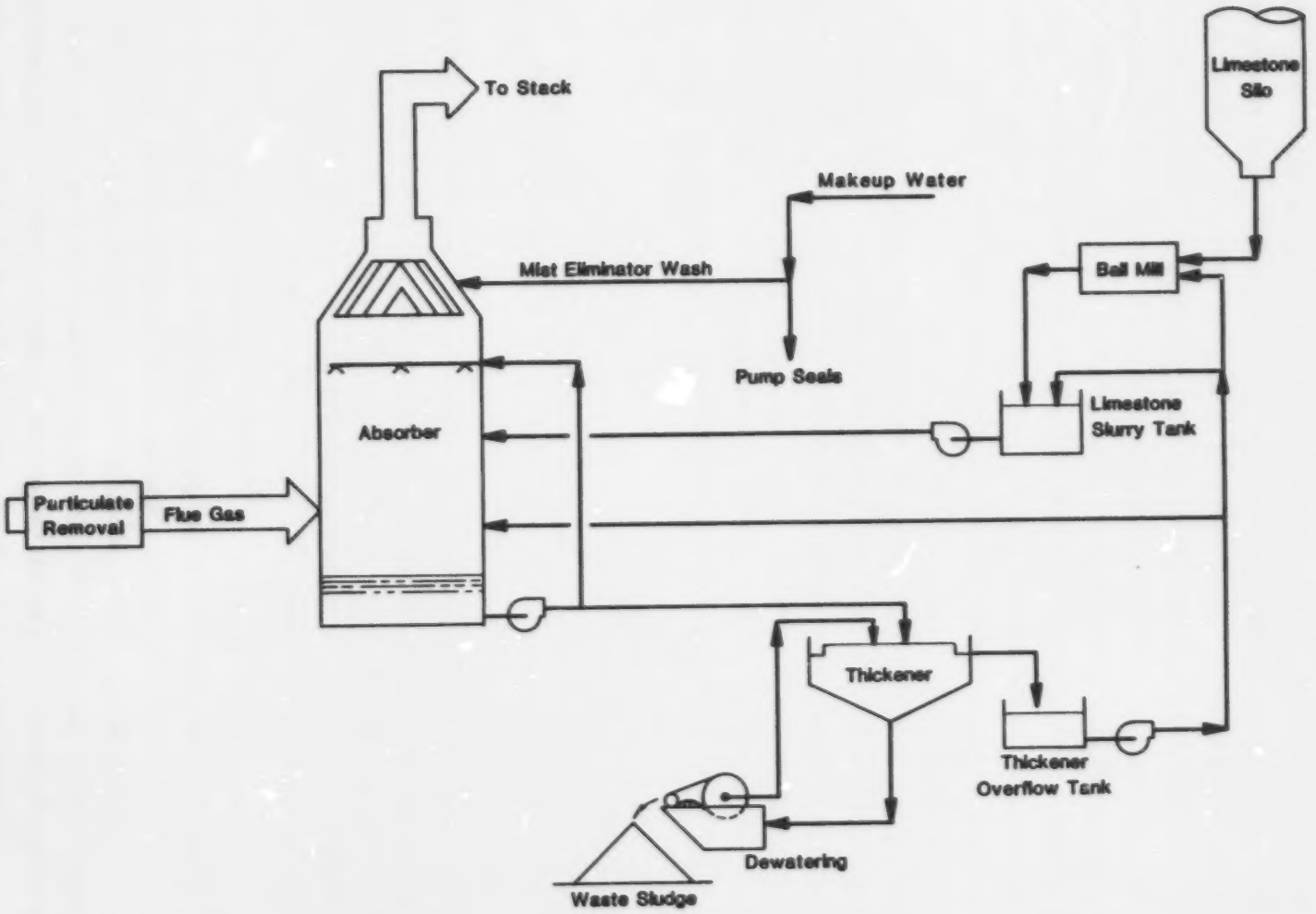


FIGURE 3 LIMESTONE SCRUBBING PROCESS SCHEMATIC

A mixed reaction tank in the absorber bottom or outside the absorber provides the necessary reaction time for the precipitation and oxidation reactions to occur. The scrubbing slurry is continuously recirculated from the reaction tank to the absorber. Makeup water and fresh reagent are added to the system, while a bleed stream is removed to maintain a solids concentration in the scrubber loop of approximately 10 to 15 percent (12). The bleed stream is typically thickened in a gravity thickener with clear overflow being returned to the scrubber loop. The thickened solids may then be dewatered further using mechanical devices such as vacuum filters or may be sent directly to disposal ponds for further thickening and dewatering. Entrained liquor released from the solids during the dewatering process is normally returned to the scrubber loop.

Scrubbed gas passes through demisters in the top of the absorber which remove liquid droplets entrained in the gas stream. As it exits the scrubber, the gas is frequently reheated to give it buoyancy and to reduce corrosion in downstream equipment. This may be accomplished using in-line reheaters, hot air injection, or partial by-pass of the hot flue gas.

Limestone scrubbing systems are typically designed to achieve SO_2 removals of up to 90 percent. Actual removals are affected by many process variables including (10,11):

a) L/G Ratios - SO_2 removal increases with increasing L/G ratios.
b) pH - Although SO_2 removal increases with increasing slurry pH, reagent utilization drops off and scaling increases, resulting in higher reagent costs and more operational problems.
c) Scrubbing Slurry Solids Concentration - SO_2 removal is generally better at higher solids concentrations. However, mechanical difficulties are encountered when handling slurries with solids concentrations much higher than 15 percent.
d) Presence of Additives - Magnesium and organic acids such as adipic acid and dibasic acid (DBA) in the scrubbing liquor enhance SO_2 removal.

The extent of oxidation of sulphite to sulphate in the scrubber liquor varies considerably between systems. Although the mechanism of sulphite oxidation is not completely understood, the oxidation rate is known to increase with increasing oxygen concentration in the flue gas and with decreasing liquor pH. It is believed that some components of fly ash may also catalyze the oxidation reaction (12). While natural oxidation results in about 25 percent conversion of sulphite to sulphate, several designs of newer systems incorporate a forced oxidation feature in which air is bubbled through the

reaction tank to achieve sulphite oxidations of 90 percent or greater. This is done to take advantage of the better dewatering and physical characteristics of sulphate sludge compared to sulphite, and in some cases to produce a marketable gypsum by-product.

While conventional limestone scrubbing systems can be modified to incorporate forced oxidation, some unique processes are available which are based on the use of this design concept. The Chiyoda Thoroughbred 121 FGD process, developed in Japan, uses a unique gas/liquid contacting device called a jet bubbling reactor (JBR) illustrated in Figure 4. Flue gas and air are sparged into the sorbent slurry in the JBR where SO_2 absorption, oxidation, and precipitation all proceed simultaneously. The advantages claimed for the process, compared to the conventional absorption tower method of gas/liquid contact, are fewer equipment requirements, the elimination of slurry recirculation, and lower energy consumption (10).

Another forced oxidation design alternative is the Bischoff process which has been used in West Germany. In this system, compressed air is injected into the absorber above the reaction tank. The air interacts with the descending low pH slurry droplets reportedly resulting in more cost-effective oxidation (10).

3.3 Lime Scrubbing Process

A typical lime scrubbing FGD process is illustrated in Figure 5. With the exception of the alkali reagent used, the lime scrubbing process configuration is virtually identical to the limestone process. Quicklime (CaO) is slaked with water according to the following reaction:

Lime Slaking: $CaO + H_2O \rightarrow Ca(OH)_2$

The absorption, neutralization, and oxidation reactions then proceed as follows:

SO_2 Absorption/Dissolution: $SO_2 + H_2O \rightarrow H_2SO_3$

Neutralization/Precipitation: $H_2SO_3 + Ca(OH)_2 \rightarrow CaSO_3 + 2H_2O$

Oxidation: $2CaSO_3 + O_2 \rightarrow 2CaSO_4$

Somewhat lower stoichiometric ratios of $Ca:SO_2$ can generally be used with lime than with limestone to achieve comparable SO_2 removals, due to the higher reactivity of the former. Nonetheless, reagent costs for lime usually exceed those for limestone systems.

Another significant difference between the two systems is their respective water quality requirements for reagent preparation. Limestone preparation is purely a mechanical wet grinding operation, for which recycled scrubber liquor is normally

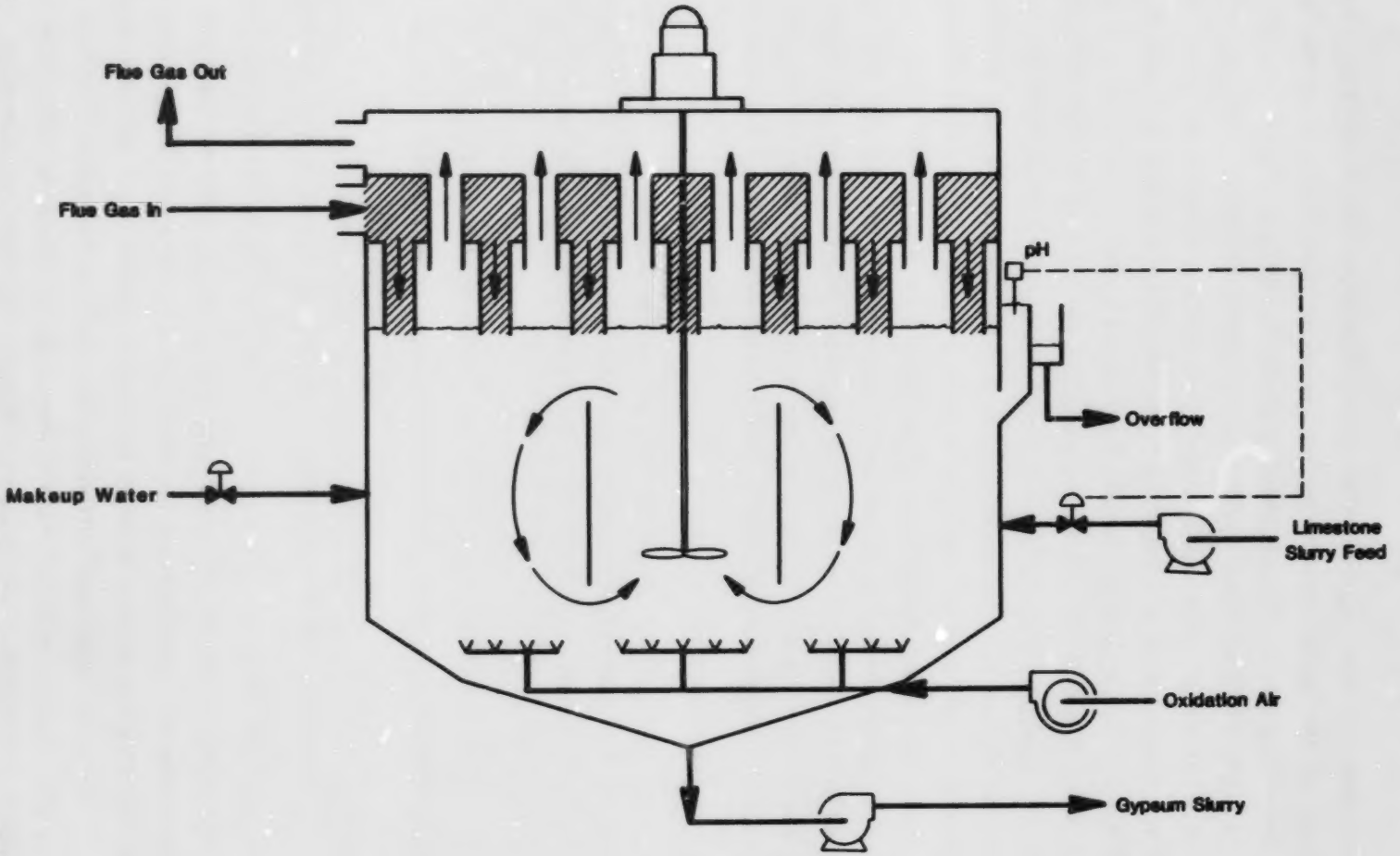


FIGURE 4 SCHEMATIC OF JET BUBBLING REACTOR (JBR)

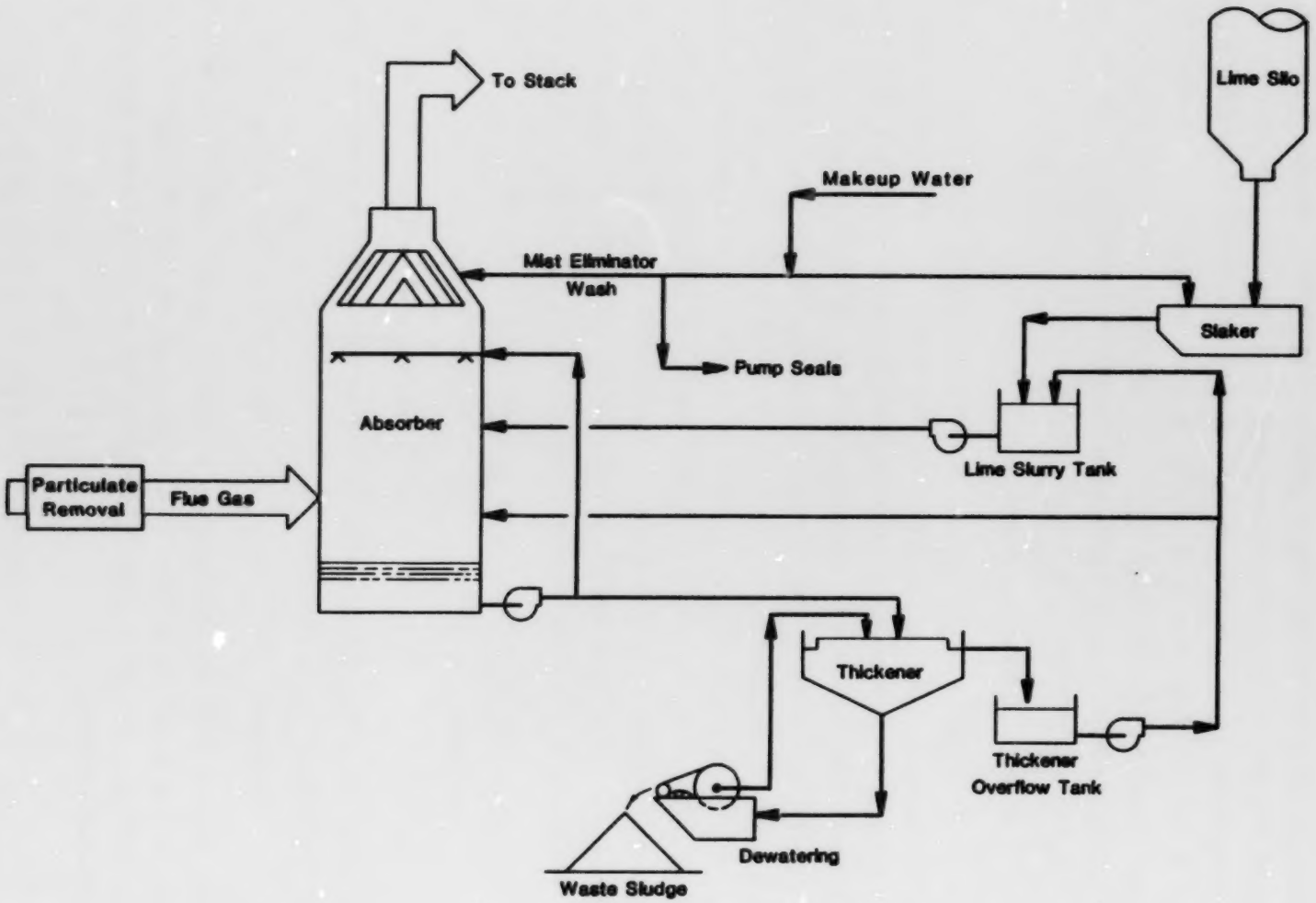


FIGURE 5 LIME SCRUBBING PROCESS SCHEMATIC

adequate for purposes of milling and slurry dilution. Lime slaking, on the other hand, is an exothermic chemical reaction which requires high quality makeup water to ensure complete reaction and good reagent reactivity in the scrubber. The presence of sulphite and sulphate ions in the slaking water, as would be encountered if scrubber liquor were reused for this purpose, retards the slaking process resulting in poor reagent utilization. This can be an important consideration in the design of closed-loop scrubbing systems where every attempt must be made to minimize fresh water inputs to the system to maintain the water balance.

3.4 Dual Alkali Scrubbing Process

The dual alkali FGD process is a two-loop system which uses a clear, sodium-based scrubbing solution for SO_2 absorption and an external calcium-based precipitation loop for reagent regeneration. Although the sodium reagent is regenerated, the process is still classified as a "throwaway" system, because the calcium salts precipitated with lime in the external loop are typically landfilled without by-product recovery. A simplified schematic of the dual alkali process is presented in Figure 6.

The dual alkali reaction chemistry is somewhat more complex than described for lime and limestone systems. In the absorption loop, SO_2 is contacted with a recirculating solution of sodium hydroxide (NaOH), sodium sulphite (Na_2SO_3) and sodium bisulphite ($NaHSO_3$), all of which are highly soluble salts. The principal SO_2 absorption reactions can be described as follows:

SO_2 Absorption/Dissolution: $SO_2 + H_2O \rightarrow H_2SO_3$

Neutralization: $2NaOH + H_2SO_3 \rightarrow Na_2SO_3 + 2H_2O$

$Na_2SO_3 + SO_2 + H_2O \rightarrow 2HaHSO_3$

Some oxidation of sodium sulphite to the slightly less soluble sulphate form also occurs according to the following reaction:

Oxidation: $2Na_2SO_3 + O_2 \rightarrow 2Na_2SO_4$

A bleed stream from the absorber loop is directed to the regeneration section, where slaked lime or limestone is added in a single stage, or multiple stage, short residence time, continuous stirred reactor. This gives rise to the following precipitation and regeneration reactions:

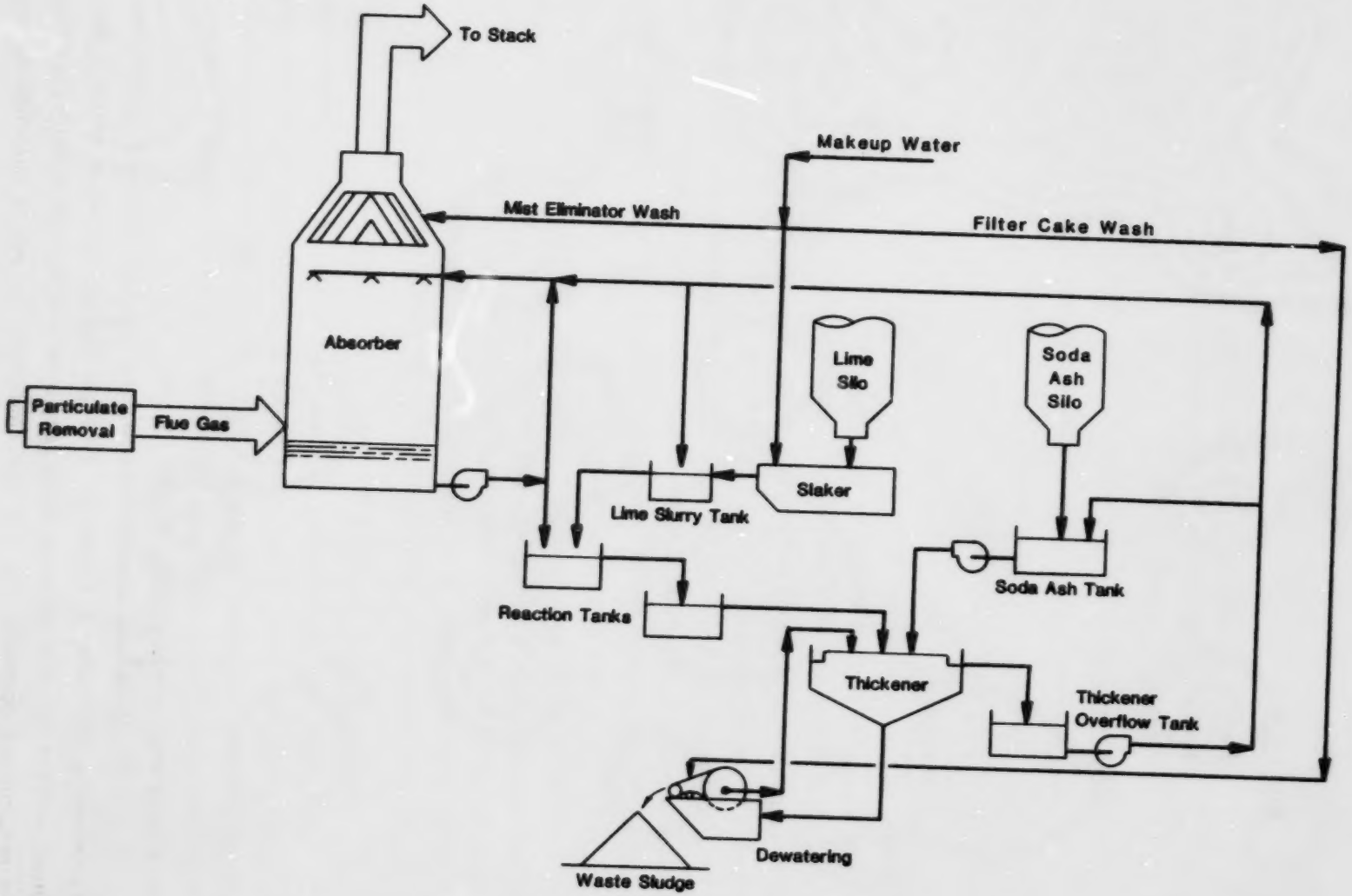


FIGURE 6 DUAL ALKALI SCRUBBING PROCESS SCHEMATIC

With Lime:

$$Na_2SO_3 + Ca(OH)_2 \rightarrow 2NaOH + CaSO_3$$
$$NaHSO_3 + Ca(OH)_2 \rightarrow NaOH + CaSO_3 + H_2O$$
$$Na_2SO_4 + Ca(OH)_2 \rightarrow 2NaOH + CaSO_4$$

With Limestone:

$$Na_2SO_3 + CaCO_3 + H_2O \rightarrow 2NaOH + CaSO_3 + CO_2$$
$$NaHSO_3 + CaCO_3 \rightarrow NaOH + CaSO_3 + CO_2$$
$$Na_2SO_4 + CaCO_3 + H_2O \rightarrow 2NaOH + CaSO_4 + CO_2$$

The resulting slurry is pumped to a clarifier or thickener, where the precipitated solids are separated from the regenerated scrubbing liquor. The underflow solids are dewatered further, usually by vacuum filtration, prior to disposal. The filter cake is washed with fresh water during the dewatering operation to minimize the loss of sodium reagent due to occlusion in the waste solids. Thickener overflow and filtrate are returned to the scrubbing loop.

Some sodium loss from the system is unavoidable, and this is usually made up by the addition of soda ash (Na_2CO_3). As soda ash addition results in a softening reaction in which calcium is precipitated as calcium carbonate, the point of addition is frequently the thickener, where the precipitated solids can be removed.

The principal advantages of the dual alkali process are that sodium is more efficient than calcium in removing SO_2, and the high solubilities of the sodium salts result in relatively scale-free absorber operation, compared to calcium-based systems. The use of solution scrubbing rather than slurry scrubbing also provides several operational advantages with respect to the severity of service for pumps, valves, etc. (10).

3.5 Alkaline Fly Ash Scrubbing Process

Alkaline fly ash scrubbing is a modified version of the lime and limestone processes. It uses the natural alkalinity of the ash to provide much of the alkali reagent necessary for SO_2 absorption and precipitation, although this is often supplemented by lime or limestone. The application of the process is generally limited to western, low-sulphur coals which produce highly alkaline fly ash.

Although several process configurations have been employed, a commonly used arrangement is illustrated in Figure 7. Flue gas leaving the boiler enters a downflow venturi scrubber in which scrubbing liquor is sprayed into the high velocity gas stream. In this prescrubbing operation the flue gas containing fly ash is quenched to near its

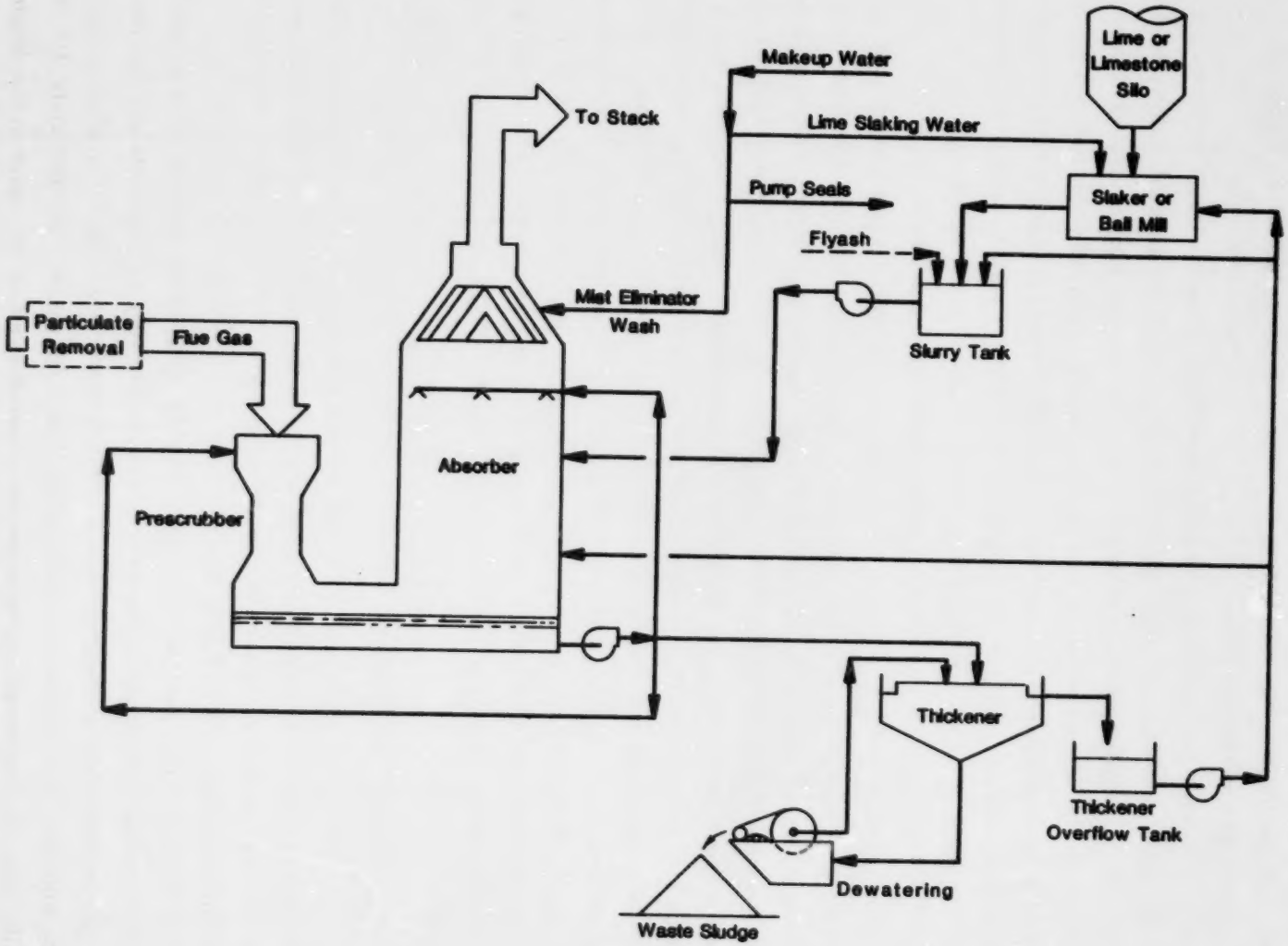


FIGURE 7 LIME/LIMESTONE ALKALINE FLY ASH SCRUBBING PROCESS SCHEMATIC

saturation temperature, particulates are wetted and captured in the falling liquid droplets, and some SO_2 absorption occurs. The gas is then directed to an upflow tower where it is contacted with circulating scrubber slurry for further SO_2 absorption and precipitation of reaction products.

As with the lime and limestone processes, a bleed stream is drawn off the scrubber slurry loop to maintain approximately a 10 to 15 percent suspended solids concentration. Alkali reagent is normally added to increase SO_2 removal or to maintain the pH within the desired operating range. The process is typically operated in a pH range of 5.0 to 5.5, as this provides good scale control and reagent utilization (the utilization of ash alkalinity increases with decreasing pH). At these pH values, extensive oxidation of calcium sulphite to sulphate frequently occurs in alkaline fly ash scrubbing systems.

The alkaline fly ash process design was employed in several early FGD systems installed on utility boilers in the western U.S.A. However, its use was limited by the availability of high alkalinity ash fuels and by an inability to meet the high SO_2 and particulate removal requirements imposed by later legislation. Recent indications suggest there is renewed interest in the use of the process in high alkaline ash applications to reduce reagent costs (10). One design modification which is being considered uses a wet electrostatic precipitator (ESP) above the absorber tower to achieve higher removals of particulates and entrained droplets than is possible with the conventional venturi/absorber design described above. Yet another modification which has been used in some systems is the removal of particulates in a dry ESP upstream of the FGD system. Captured fly ash is then slurried with scrubber liquor and fed to the process as reagent. The advantages of this configuration are higher particulate removal efficiencies and better process control in the FGD system.

3.6 Spray Dryer Scrubbing Process

The spray dryer, dry scrubbing process has captured a high degree of interest in the utility industry in recent years due largely to the simplicity of the system and the superior handling characteristics of the dry solid waste that it produces. The process is illustrated schematically in Figure 8.

Boiler flue gas passes directly into the spray dryer without pretreatment for fly ash removal. The scrubbing liquor is atomized into very fine droplets and sprayed into the gas stream. SO_2 is absorbed into the droplets and reacts with the alkali reagent to form sulphites and sulphates. Sufficiently low L/G ratios (approximately 100 times smaller than for wet scrubbers) are used to ensure that all of the water in the droplets is

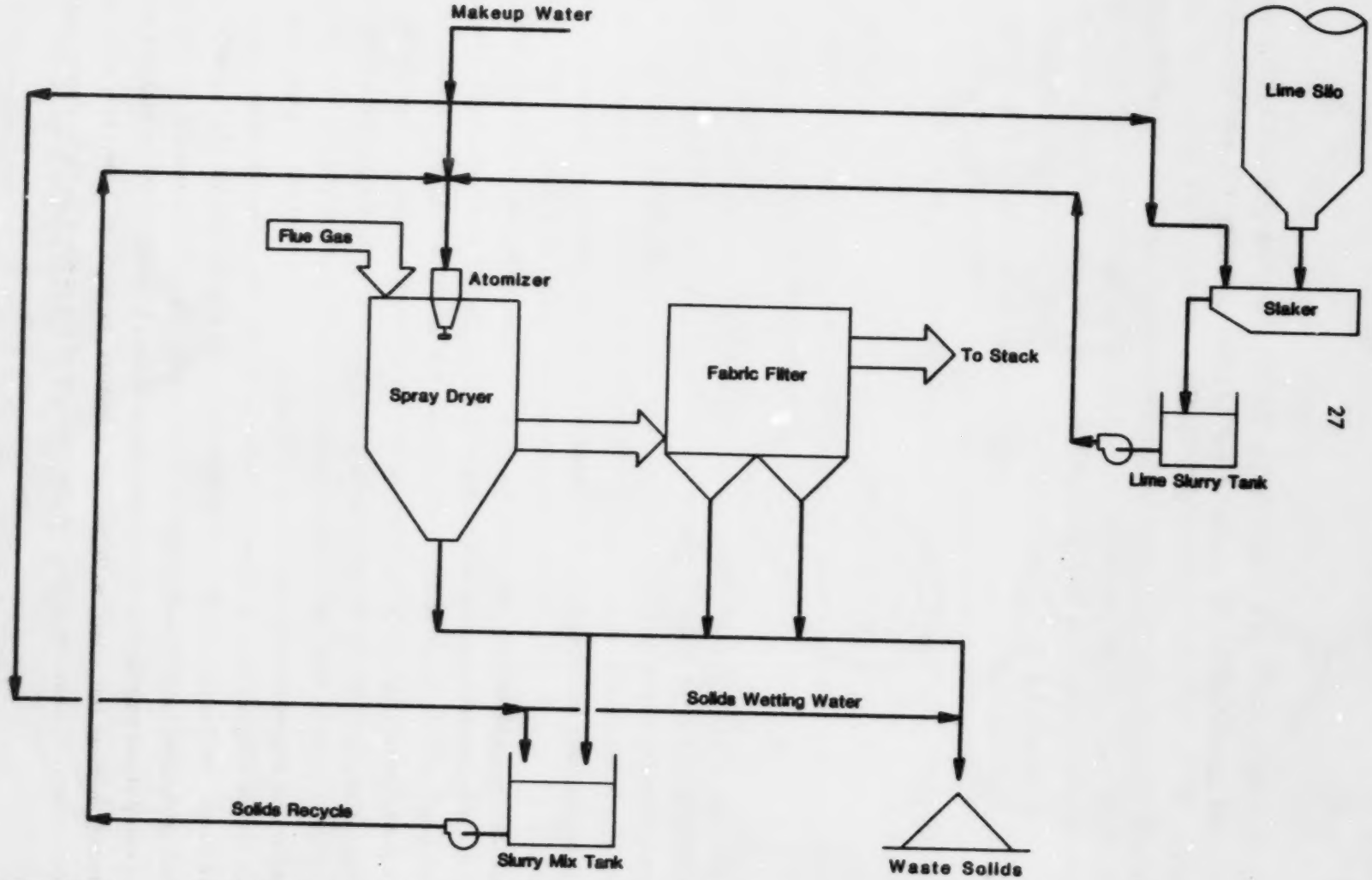


FIGURE 8 SPRAY DRYER SCRUBBING PROCESS SCHEMATIC

vapourized (10). With some systems, a portion of the dry reaction products, fly ash, and unreacted sorbent fall to the bottom of the spray dryer, where they are removed, while the rest of the solids are captured in a downstream fabric filter (baghouse). In other systems, all particulate removal is accomplished in the baghouse. Unreacted sorbent collects on the fabric as a filter cake, providing significant SO_2 removal in the baghouse due to additional absorption and reaction as the flue gas passes through.

Both lime and soda ash have been used as scrubbing reagents. If lime is used, the reaction chemistry is identical to that described for the lime wet scrubbing process and the products are calcium sulphite and sulphate. When soda ash is used, sodium sulphite and sulphate are produced as described by the following reactions:

SO_2 Absorption/Dissolution:	$SO_2 + H_2O \rightarrow H_2SO_3$
Neutralization:	$Na_2CO_3 + H_2SO_3 \rightarrow Na_2SO_3 + CO_2 + H_2O$
Oxidation:	$2Na_2SO_3 + O_2 \rightarrow 2Na_2SO_4$

While soda ash gives higher SO_2 removals than lime at similar stoichiometric ratios, it is much more expensive to use.

Most spray dryer systems sold to date have been for medium to low sulphur coals (generally less than 1.5 percent sulphur). In early tests, the process demonstrated an ability to achieve SO_2 removal efficiencies of 75 to 85 percent with high reagent utilization. However, high stoichiometric ratios with corresponding low reagent utilization were generally required to achieve greater than about 85 percent SO_2 removal (13). As SO_2 removal requirements of 90 percent or greater would normally be required with high sulphur coals, spray drying was considered uneconomic in this application because of the high reagent costs associated with such removals.

More recently, tests with recycling of the waste solids captured in the spray dryer and baghouse have shown some promising results with improved reagent utilization at higher SO_2 removal efficiencies (14). With this design modification, the solids are reslurried with water and fed to the spray dryer with fresh reagent. This enables reuse of the unreacted sorbent and available alkali in the fly ash, decreasing the fresh sorbent makeup requirements. Vendor guarantees for SO_2 removals of 90 percent are reportedly being offered for coals with sulphur contents as high as 3 percent, although some controversy remains regarding the applicability of the process to high sulphur coals (10).

The spray drying process control is based on two primary loops. Outlet temperature controls the water feed rate to the atomizer, while outlet SO_2 concentration controls the reagent concentration. The process normally operates at an outlet temperature at least 10°C above the adiabatic saturation temperature. Although SO_2 removal efficiency increases significantly at lower temperatures, so does the potential for accumulation of wet solids in the dryer and the particulate collector, which creates severe operating problems. Because the system operates well above the dew point of the gas, no reheat is required prior to discharge.

Efficient atomization of the reagent is widely recognized as the key to the success of the spray drying operation. Both rotary and nozzle designs are being promoted by vendors. The rotary design is claimed to offer better control over droplet size, lower energy consumption and less potential for erosion, scaling and plugging, while the nozzle is less expensive, mechanically less complex and easier to maintain.

3.7 Wellman-Lord Process

Of the regenerative FGD processes, the Wellman-Lord is likely the best known. It uses a concentrated sodium sulphite scrubbing solution to absorb SO_2 and produce a sulphite/bisulphite liquor. The liquor is thermally treated in evaporators to regenerate the sodium sulphite scrubbing solution and to drive off a concentrated stream of SO_2 which can then be further processed to produce elemental sulphur, sulphuric acid, or liquid SO_2. The process is illustrated schematically in Figure 9.

Flue gas is generally pretreated for particulate removal in an electrostatic precipitator or fabric filter. It is then passed through a prescrubber, where water is used to cool and humidify the gas, and to remove residual fly ash and chlorides. The upstream removal of chlorides reduces the potential for stress corrosion in downstream equipment, and controls the buildup of dissolved solids. Chlorides, like other tramp anions in the scrubbing liquor, tie up sodium cations necessitating higher reagent makeup rates to maintain SO_2 removal efficiency (10). Blowdown from the prescrubber loop is normally sent to a disposal pond to control the buildup of suspended solids and chlorides.

Cooled and humidified gas from the prescrubber passes upward through the absorption tower, countercurrent to the flow of sodium sulphite scrubbing solution. SO_2 is absorbed and reacts to form the more soluble sodium bisulphite according to the following reactions:

SO_2 Absorption/Dissolution: $SO_2 + H_2O \rightarrow H_2SO_3$

Neutralization: $Na_2SO_3 + H_2SO_3 \rightarrow 2NaHSO_3$

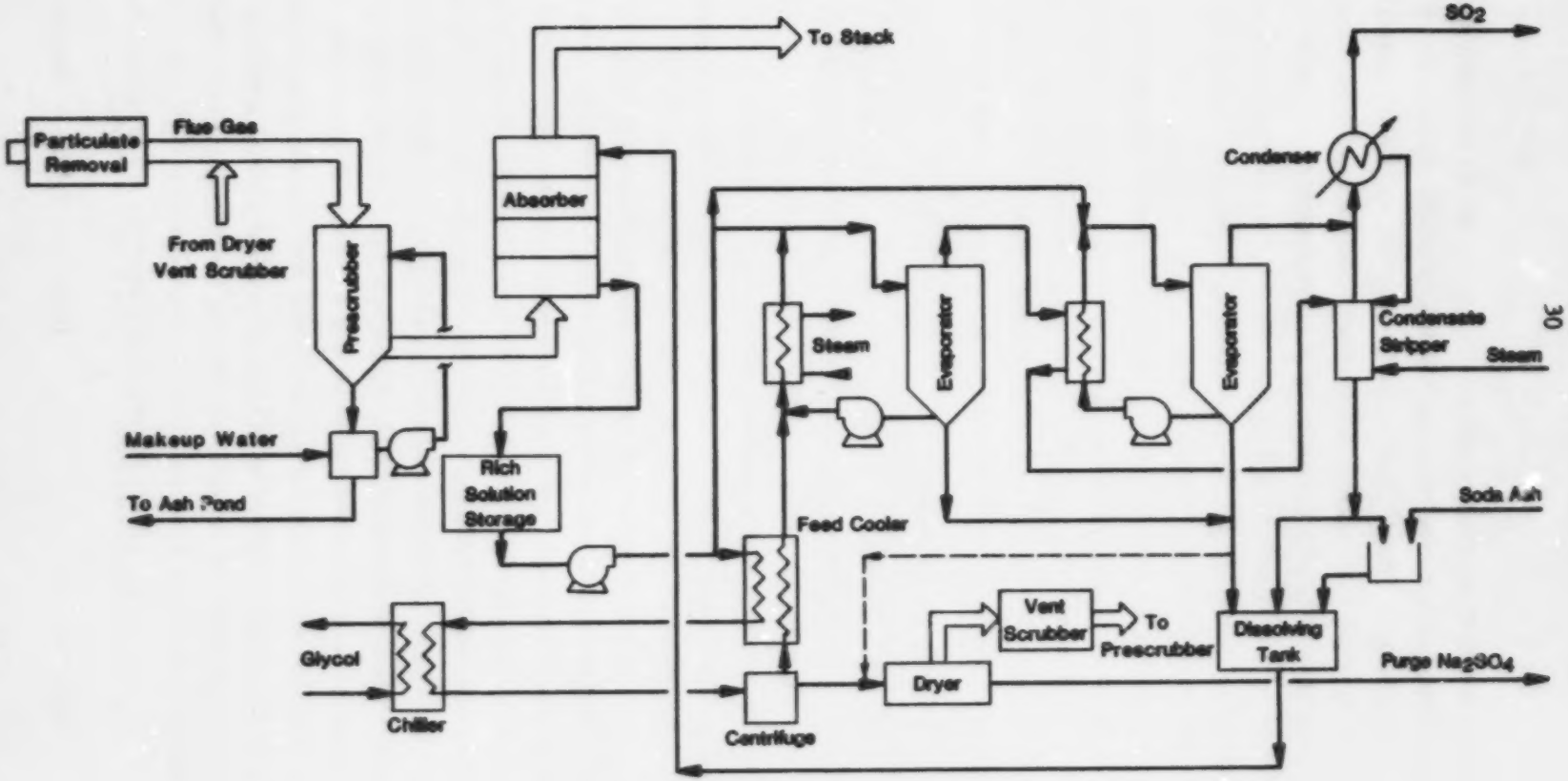


FIGURE 9 WELLMAN-LORD PROCESS SCHEMATIC

Make-up soda ash in the scrubbing liquor also reacts to form sodium sulphite and bisulphite as follows:

$$Na_2CO_3 + H_2SO_3 \rightarrow Na_2SO_3 + CO_2 + H_2O$$
$$Na_2CO_3 + 2H_2SO_3 \rightarrow 2NaHSO_3 + CO_2 + H_2O$$

Some sodium sulphate is also formed by oxidation:

$$2Na_2SO_3 + O_2 \rightarrow 2Na_2SO_4$$

The bisulphite-rich scrubbing solution is discharged from the bottom of the absorber into a surge tank. A purge stream is routed to the purge treatment area for sodium sulphate removal, while the rest is sent to an evaporator for sodium sulphite regeneration.

Regeneration of sodium sulphite is accomplished by adding heat, which reverses the absorption/neutralization reactions as follows:

$$2NaHSO_3 \rightarrow Na_2SO_3 + H_2O + SO_2$$

The regeneration reaction is limited by the equilibrium concentration of sulphite ion in solution. However, since sodium sulphite is less soluble than sodium bisulphite, it is continuously removed from solution by crystallization, thus driving the reaction forward.

Sodium thiosulphate ($Na_2S_2O_3$) also forms in the evaporator as follows:

$$6NaHSO_3 \rightarrow 2Na_2SO_4 + Na_2S_2O_3 + 2SO_2 + 3H_2O$$
$$2NaHSO_3 + 2Na_2SO_3 \rightarrow 2Na_2SO_4 + Na_2S_2O_3 + H_2O$$

The feed stream to the evaporator is usually split between two effects, with approximately 55 percent going to the first effect, and 45 percent to the second. The first effect operates at about 95°C and is heated with low-pressure steam by an external shell and tube heat exchanger. The SO_2 and water vapour driven off overhead from the first effect are used to heat the second effect which operates at about 75°C (12).

The overhead SO_2/water vapour mixture from the evaporators is passed through a series of partial condensers to remove most of the water and concentrate the SO_2. The condensate, containing several hundred parts per million dissolved SO_2, is steamstripped to reduce this value and is then sent to a dissolving tank, where crystallized sodium sulphite from the evaporators and makeup soda ash are dissolved to form fresh scrubbing solution.

A sidestream of the absorber scrubbing liquor is drawn off prior to the evaporator for removal of sodium sulphate in a chiller/crystallizer. This stream is

precooled by heat exchange with cold, sulphate-free centrate. When the purge is cooled to about 0°C in the chiller/crystallizer, a mixture of sodium sulphate and sulphite crystallizes. The resulting slurry is centrifuged to produce a cake containing about 40 percent solids (12). Prior to steam drying of the cake a secondary purge stream is drawn off the evaporators and added to the cake to remove the thiosulphate formed in regeneration. The resulting product is a crystalline mixture of approximately 70 percent anhydrous sodium sulphate and 30 percent sodium sulphite, with small amounts of thiosulphates, pyrosulphates, and chlorides (12). It is frequently sold as reagent to the kraft paper industry. The sulphate-free supernatant liquor from the centrifuge is heated by passing it through the feed cooler, and then returned to the product liquor stream via the evaporator loop. The vent gases from the dryer are cleaned to remove dust and are returned to the main flue gas stream ahead of the prescrubber.

The Wellman-Lord process has been widely used in the utility industry in Japan and has also been used to scrub tail gases from sulphuric acid plants and oil refineries. The process has seen limited application to coal-fired utility boilers in North America. However, all commercial units in the U.S.A. and abroad reportedly achieve better than 90 percent SO_2 removal (10).

3.8 Other Processes

Although the six FGD processes described above are the most commonly used in the North American utility industry at present, several others have been used in the U.S.A. and other countries. While some of these are unlikely to be applied in Canada, others may be considered potential candidates by electrical utilities. Seven such processes are briefly described below.

3.8.1 Sodium Carbonate Scrubbing Process. The sodium carbonate scrubbing process is essentially a once-through, "throwaway" system using trona (naturally-occurring sodium carbonate) as the sorbent. Spent liquor from the absorber is pumped to settling ponds, where the water is allowed to evaporate. No attempt is made to regenerate the scrubbing reagent or to recover a by-product.

The process is only practical in arid regions where the natural evaporation rate exceeds the combined rate of spent liquor production and rainfall. The highly soluble sodium sulphite and bisulphite compounds in the waste liquor preclude discharge to surface waters. The high sodium makeup requirements for this once-through process also make it prohibitively expensive unless an inexpensive source of the alkali reagent used is available.

For these reasons, most utilities that have used this process are located in the western U.S.A., where large tracts of land are available for ponding, the evaporation rate is high, low sulphur coal is used, and naturally-occurring deposits of the reagent are locally available.

3.8.2 Seawater Scrubbing Process. A unique once-through type FGD process which has been used in Norway is the seawater scrubbing process. The process uses the natural bicarbonate (HCO_3^-) alkalinity of seawater to absorb and react with SO_2 as follows:

$$SO_2 + H_2O \rightarrow 2H^+ + SO_3^=$$

$$HCO_3^- + H^+ \rightarrow CO_2 + H_2O$$

The process is designed for application at stations using seawater for condenser cooling. Water leaving the condensers is fed to a countercurrent absorption tower where SO_2 is absorbed from the flue gas. The scrubber is operated in strictly a once-through manner, with no recirculation of liquor.

In recent designs, the acidic effluent from the absorber is sent to a treatment process where additional seawater from the condensers and lime are added for neutralization, and air is supplied to oxidize the sulphite to sulphate. Sufficient seawater is added to avoid the precipitation of remaining calcium sulphite, which is less soluble than calcium sulphate.

The gas temperature at the scrubber outlet is close to the seawater inlet temperature. This is much lower than for conventional wet scrubbers. In fact, the cooling of the gas results in the condensation of water vapour from the combustion products, rather than the evaporation of scrubber liquor as occurs in conventional systems (15). Flue gas by-pass is normally used for reheating.

The advantages claimed for the process are operation in a solution scrubbing mode (which avoids scaling, plugging, and erosion problems associated with slurry scrubbers), low chemical consumption (compared to conventional lime or limestone scrubbers), simplicity of design and low capital cost (due to the lack of solid waste handling and disposal). The principal disadvantages associated with the process identified by environmental regulatory agencies are the comparatively high water use and the waste discharges associated with operation in a once-through mode, both of which are contrary to the intent of the proposed Environmental Code of Practice for the Canadian Steam-Electric Power Generation Industry (16).

3.8.3 Davy-Saarberg Hoelter Process. The Davy-Saarberg Hoelter (Davy S-H) FGD process developed in West Germany uses a clear, alkaline scrubbing solution and produces a gypsum by-product. It differs from conventional lime and limestone forced oxidation processes in that it uses solution rather than slurry scrubbing.

Flue gas is scrubbed cocurrently in an absorption tower using a clear solution of calcium hydroxide ($Ca(OH)_2$), calcium formate ($Ca(COOH)_2$), and calcium chloride ($CaCl_2$). A water-soluble intermediate, calcium bisulphite ($Ca(HSO_3)_2$), is formed, which is then oxidized to calcium sulphate in a reaction tank, precipitated, and recovered as gypsum following solid/liquid separation.

Calcium hydroxide, or hydrated lime, the active component of the scrubbing solution, is fed to the absorber at a pH of about 9.5 to 10.5. It reacts with absorbed SO_2 to form calcium bisulphite as follows:

SO_2 Absorption/Dissolution: $SO_2 + H_2O \rightarrow H_2SO_3$

Neutralization: $Ca(OH)_2 + 2H_2SO_3 \rightarrow Ca(HSO_3)_2 + 2H_2O$

The consumption of solution alkalinity in the above reaction causes the pH to drop rapidly to about 5. However, the presence of formic acid buffers the solution to the pH 4.5 to 5 range, which reportedly assures the formation of calcium bisulphite (17).

Air is injected to a reaction tank at the bottom of the absorber to convert calcium bisulphite to calcium sulphate. Ferric chloride is added to catalyze the reaction, which proceeds at a pH of about 3.5 to 4.0 (17). The calcium sulphate slurry overflows to a thickener where hydrated lime is added to replenish calcium ions precipitated as gypsum and to increase the pH back to the 9.5 to 10.5 range used for scrubbing. Thickener underflow is dewatered further by vacuum filtration to produce a dry (75 to 80 percent solids) gypsum filter cake. Filtrate is recirculated to the thickener, and thickener overflow is returned to the absorber as fresh scrubbing solution. Formic acid is added to replenish losses due to occlusion in the filter cake. If the flue gas does not contain chlorides, hydrochloric acid is also added to increase the chloride concentration in the scrubbing liquor. This reportedly increases the solubility of calcium hydroxide, thereby improving SO_2 removal efficiency and reagent utilization.

By using solution scrubbing, the process offers the operational advantages of fewer scaling, plugging and erosion problems than those encountered with conventional forced oxidation, slurry systems. However, process control is somewhat more complex.

3.8.4 Dry Injection. One dry scrubbing FGD process has been demonstrated which differs from spray drying in that it uses absolutely no water. Dry powdered sorbent is

injected directly into the flue gas ductwork upstream of a fabric filter. The sorbent collects on the fabric, and SO_2 is removed as the flue gas passes through the filter cake. The most suitable sorbents used to date have been trona (naturally-occurring sodium carbonate) and nahcolite (naturally-occurring sodium bicarbonate).

In one demonstration test on a 22-MW coal-fired boiler, SO_2 removals of 75 to 83 percent were observed with the injection of a stoichiometric quantity of nahcolite. SO_2 removal was found to be strongly dependent on operating temperature, as removals dropped to about 40 percent at similar stoichiometric ratios when the baghouse temperature decreased from 150°C to 135°C (18).

The advantages of the process include simplicity of design, elimination of all liquid handling systems, minimal equipment requirements, and low capital cost. However, its application is likely to be limited to low sulphur coals with less stringent SO_2 removal requirements because of the relatively high stoichiometric ratios of reagent required to achieve removals much in excess of 80 percent. The availability of reagents will likely further limit the use of the process to certain regions.

3.8.5 Magnesium Oxide Process. The magnesium oxide (MgO) scrubbing process is a regenerable FGD process in which hydrated magnesium oxide absorbs and reacts with SO_2 to form magnesium sulphite and sulphate crystals. The solids are dewatered, dried and calcined to roast off SO_2 and regenerate MgO reagent. The SO_2 is then further processed to sulphuric acid or elemental sulphur.

Flue gas is pretreated in a wet scrubber for removal of chlorides and particulates. It is then directed to the absorber where SO_2 absorption and reaction occurs to produce magnesium sulphite as follows:

Slaking: $MgO + H_2O \rightarrow Mg(OH)_2$

SO_2 Absorption/Dissolution: $SO_2 + H_2O \rightarrow H_2SO_3$

Neutralization/Precipitation: $Mg(OH)_2 + H_2SO_3 \rightarrow MgSO_3 + 2H_2O$

Some oxidation of sulphite to sulphate also occurs:

$$2MgSO_3 + O_2 \rightarrow 2MgSO_4$$

The scrubbing slurry is recirculated to the absorber, while a bleed stream is sent to a thickener to maintain about a 10 - 15 percent solids concentration in the recirculating liquor. Thickener overflow is returned to the scrubber loop while the underflow is centrifuged to recover a wet cake of hydrated $MgSO_3$ and $MgSO_4$ crystals.

The wet crystals are discharged from the centrifuge into a rotary or fluid-bed dryer where both the free and chemically bound water of hydration are removed. The dried solids are conveyed from the dryer to a calciner for MgO regeneration and SO_2 recovery. The dried $MgSO_3/MgSO_4$ crystals are calcined in a rotary kiln or fluidized-bed reactor. The thermal decomposition reaction of $MgSO_3$ is shown below:

$$MgSO_3 \rightarrow MgO + SO_2$$

$MgSO_4$ is reduced in the presence of carbon (coke) as follows:

$$2MgSO_4 + C \rightarrow 2MgO + 2SO_2 + CO_2$$

The SO_2-rich off-gas from the calciner is sent to either a sulphur recovery or sulphuric acid production unit, while regenerated MgO is slaked with water and returned to the scrubbing loop.

The magnesium oxide process offers several advantages over calcium and sodium based systems:

a) Higher reagent solubilities mean more efficient SO_2 removal at lower L/G ratios.
b) The higher solubility of magnesium sulphite and sulphate compared to the calcium salts reduces scaling problems.
c) The sulphite and sulphate crystals can both be regenerated to MgO, eliminating the need for a purge from the system.

724 MW of MgO scrubbing capacity is scheduled for startup in 1983. This will represent the first commercial demonstration of the process in the North American utility industry (9).

3.8.6 Aqueous Carbonate Process. The aqueous carbonate process is a regenerative FGD system that uses a spray dryer scrubbing technique for SO_2 removal. A solution of sodium carbonate absorbs SO_2 in a spray dryer. The dry product is then thermally regenerated to sodium carbonate and hydrogen sulphide (H_2S). The H_2S can then be further processed to sulphur by-products.

The process differs from conventional spray drying in that the flue gas is pretreated for fly ash removal to prevent reagent contamination during regeneration. The precleaned gas is directed to a spray dryer where sodium carbonate scrubbing solution is introduced as a fine mist to form a dry sodium sulphite and sulphate product. The absorption, neutralization and oxidation reactions are described as follows:

SO_2 Absorption/Dissolution: $SO_2 + H_2O \rightarrow H_2SO_3$

Neutralization: $Na_2CO_3 + H_2SO_3 \rightarrow Na_2SO_3 + CO_2 + H_2O$

Oxidation: $2Na_2SO_3 + O_2 \rightarrow 2Na_2SO_4$

Some of the dry, powdered reaction products and unreacted reagent are separated from the flue gas within the spray dryer, while the balance is removed in downstream particulate removal equipment. The captured solids are then transferred to the regeneration loop where they are heated with coke or another carbon source to approximately 930 to 1040°C. This reduces the sodium sulphite and sulphate to sodium sulphide (Na_2S) and carbon dioxide. The molten Na_2S flows to a quench tank, where it is dissolved in water. The aqueous solution is then filtered for removal of residual coke, coke ash, and fly ash. After filtration, the solution is recarbonated with CO_2 to produce hydrogen sulphide off-gas and regenerated sodium carbonate scrubbing solution.

The regenerated reagent is returned to the spray dryer, while the H_2S is sent to a conversion facility, such as a Claus plant, for production of by-product sulphur. The regeneration process is described by the following reactions:

Reduction: $2Na_2SO_3 + 3C \rightarrow 2Na_2S + 3CO_2$

$Na_2SO_4 + C \rightarrow Na_2S + 2CO_2$

Recarbonation: $Na_2S + H_2O + CO_2 \rightarrow Na_2CO_3 + H_2S$

The first 100-MW commercial demonstration plant is currently in the early stages of operation in the U.S.A. (9).

3.8.7 Citrate Process. The citrate process is a regenerable FGD process in which elemental sulphur is precipitated directly from the scrubber solution. SO_2 is absorbed in a sodium citrate scrubbing solution which is then reacted with gaseous H_2S to precipitate elemental sulphur and regenerate the scrubbing liquor. Sulphur is separated from the liquor by flotation and approximately two-thirds of the by-product sulphur is reused to produce the H_2S needed for regeneration.

Boiler flue gas must be precleaned for fly ash removal to avoid contamination of the product sulphur. It is then directed to an absorption tower where a solution of sodium citrate, citric acid, and sodium thiosulphate absorb and react with the SO_2 to form soluble sodium bisulphite. The citrate ion buffers the scrubbing solution at a pH of about 3.0 to 3.7, which apparently suppresses side reactions. The absorption reactions are described as follows:

SO_2 Absorption/Dissolution: $SO_2 + H_2O \rightarrow H_2SO_3$

Reaction: H_2SO_3 + Na•citrate → $NaHSO_3$ + citric acid

The SO_2-rich scrubbing solution is pumped from the absorber to a regeneration reactor, where H_2S is introduced to precipitate elemental sulphur as follows:

$NaHSO_3$ + citric acid + $2H_2S \rightarrow 3S + 3H_2O$ + Na•citrate

Off-gas from the reactor, which is mainly CO_2 with small quantities of H_2S, flows to the boiler, where it is incinerated to SO_2. Sulphur is separated from the regenerated sodium citrate solution by air flotation, and is removed as a froth containing about 10 percent solids. This slurry is then heated to about 130°C to melt the sulphur which is gravity separated from the citrate solution under pressure, and filtered to remove solid impurities. Approximately two-thirds of the product sulphur is reacted with steam and natural gas in a catalytic reactor to produce the H_2S used in regeneration, according to the following reaction:

$CH_4 + 4S + 2H_2O \rightarrow 4H_2S + CO_2$

Some oxidation occurs in the absorber loop, resulting in the formation of sulphates. These are purged from the liquor by chilling a bleed stream of the citrate solution and crystallizing Glauber's Salt ($Na_2SO_4 \cdot 10H_2O$) which is removed by filtration and sold as by-product or landfilled.

The first North American demonstration of the process was a 60-MW unit commissioned in 1980. However, the system has experienced numerous operating problems, resulting in continuous runs to date of only seven to ten days (10).

4 FGD-EQUIPPED STATIONS VISITED

Visits were conducted to several electric power generating stations in the U.S.A. which had demonstrated a significant history of experience with FGD systems considered potentially applicable to Canadian coals. The objectives of the visits were to compile design, performance and cost data on the various FGD systems; to discuss the operational experiences, problems and successes with station engineering and operating personnel; and to collect samples of FGD wastes at selected stations to perform waste characterization studies.

In Phase One of this study, six stations burning medium to high sulphur coals and using lime, limestone, dual-alkali, and Wellman-Lord FGD systems were visited. A summary of stations visited in Phase One is presented in Table 8, and their principal operating characteristics are outlined in Table 9 (2).

Since several of the processes examined in Phase One were considered applicable to the higher sulphur coals typically burned in eastern Canada, the Phase Two site visits concentrated on facilities using processes which would be applicable to the lower sulphur lignite and subbituminous coals characteristic of western Canada. Stations using alkaline fly ash and spray dry FGD systems comprised the majority of the visits. However, one station which had pilot tested the Chiyoda Thoroughbred 121 process was also visited because of interest in the unique features and apparent advantages offered by this process over conventionally-designed limestone FGD systems.

A summary of stations visited in Phase Two is presented in Table 10, and their principal operating characteristics are outlined in Table 11. Process details for each system are presented in Appendix A.

TABLE 8 SUMMARY OF STATIONS VISITED IN PHASE ONE

Utility	Station/Unit	Capacity (MW)	Coal Type		FGD Process
			Heating Value (MJ/kg)	Sulphur (%)	
Central Illinois Public Service, Newton, Illinois	Newton 1	590	27.1	3.00	Dual Alkali
Louisville Gas & Electric, Louisville, Kentucky	Cane Run 4	188	25.6	3.75	Lime
	Cane Run 5	200	25.6	3.75	Lime
	Cane Run 6	299	25.6	3.75	Dual Alkali
Northern Indiana Public Service, Gary, Indiana	Dean H. Mitchell 11	115	24.4	3.25	Wellman-Lord
Pennsylvania Power, Shippingport, Pennsylvania	Bruce Mansfield 1	917	26.7	3.00	Lime
	Bruce Mansfield 2	917	26.7	3.00	Lime
	Bruce Mansfield 3	917	26.7	3.00	Lime
Southern Illinois Power, Marion, Illinois	Marion 4	173	23.3	3.50	Limestone
Springfield City Water Light & Power, Springfield, Illinois	Dallman 3	205	24.4	3.30	Limestone with Forced Oxidation

TABLE 9 SUMMARY OF OPERATING CHARACTERISTICS OF STATIONS VISITED IN PHASE ONE

Station	FGD Process	Equipment Housing	% SO_3/% SO_4	Waste Management: Sludge Dewatering	Sludge Fixation/ Stabilization	Sludge Disposal	Monitoring Program	Wastewater Discharge	Major Problems
Newton	Dual Alkali	Reagent prep. & sludge dewatering & fixation equipment	85/15	Thickened & vacuum filtered to 60% solids	Blended with fly ash & lime to 75-80% solids (IUCS)	Landfill in natural clay site	Groundwater & runoff at landfill site monitored	None	Materials problems (corrosion) Sludge dewatering
Cane Run 4 & 5	Lime	Reagent prep. & sludge dewatering & fixation equipment	90/10	Thickened & vacuum filtered to 50-55% solids	Blended with fly ash & lime (IUCS)	Landfill in clay-lined site	Groundwater & runoff monitored at landfill site	Occasional	Materials problems (corrosion)
Cane Run 6	Dual Alkali	Reagent prep. & sludge dewatering & fixation equipment	85/15	Thickened & vacuum filtered to 50-55% solids	Blended with fly ash & lime (IUCS)	Landfill in clay-lined site	Groundwater & runoff monitored at landfill site	None	Absorber tray pluggage Sludge dewatering
Dean H. Mitchell	Wellman-Lord	Reagent prep., purge solids & SO_2 reduction equipment	N/A	Centrifuged & dried	N/A	Sulphur sold Na_2SO_4 sold or sent to bottom ash pond	None	Prescrubber blowdown to once-through fly ash pond	Lack of redundancy Inadequate capacity
Bruce Mansfield	Lime	All process equipment	-	Thickened to 20% solids	Blended with fly ash & calcilox	Unlined pond	Groundwater monitoring around disposal pond	Continuous from disposal pond	Materials problems (corrosion) Instrumentation problems Line plugging
Marion	Limestone	Reagent prep. & sludge dewatering & stabilization equipment	90/10	Thickened to 30% solids Centrifuged to 60% solids	Blended with fly ash & bottom ash	Landfill in native silt (unlined) site	None	Intermittent	Equipment freezing Equpment abrasion Materials corrosion
Dallman	Limestone with forced oxidation	All process equipment	25/75	Thickened in hydrocyclone to 55% solids and vacuum filtered to 80% solids	Forced oxidation	Landfill in natural clay site	Groundwater monitoring at landfill site	None	Materials problems (corrosion & abrasion) Insufficient air flow for complete oxidation

N/A - Not Applicable

TABLE 10 SUMMARY OF STATIONS VISITED IN PHASE TWO

Utility	Station/Unit	Capacity (MW)	Coal Type		FGD Process
			Heating Value (MJ/kg)	Sulphur (%)	
Northern States Power Co., Becker, Minnesota	Sherburne 1&2	1440	20.3	0.85	Limestone-Alkaline Fly Ash
Minnkota Power Cooperative Inc. Center, North Dakota	Milton R. Young 2	440	15	0.7	Lime-Alkaline Fly Ash
Montana - Dakota Utilities Co., Beulah, North Dakota	Coyote 1	440	15.8	0.65	Soda Ash - Spray Drying
Northern States Power Co., Minneapolis, Minnesota	Riverside 6&7	110	21.6	1.2	Lime - Spray Drying
Gulf Power Co., Chattahoochee, Florida	Scholz 1&2	80	29.3	2.0	Chiyoda* Thoroughbred 121 (CT-121)

* CT-121 process operated at Scholtz station was a 23-MW pilot unit.

TABLE 11 SUMMARY OF OPERATING CHARACTERISTICS OF STATIONS VISITED IN PHASE TWO

Station	FGD Process	System Housing	% SO_3/% SO_4	Waste Characterization and Management					Major Problems
				Dewatering	Fixation/ Stabilization	Sludge Disposal	Monitoring Program	Wastewater Discharge	
Sherburne	Limestone-Alkaline Fly Ash	All process equipment except thickeners	2/98	Thickener & pond	None	Sludge Pond	Groundwater monitoring around disposal pond	None	Plugging of absorber internals & mist eliminators Materials corrosion & erosion
Milton R. Young	Lime-Alkaline Fly Ash	Reagent prep. & sludge dewatering equipment	0/100	Pond (thickener & vacuum filter abandoned)	None	Minefill	Groundwater monitoring around disposal pond	None	Demister plugging Lining failures in absorbers Piping erosion Freezing of exposed lines & equipment
Coyote	Soda Ash Spray Drying	All process equipment	12/88	N/A	Dry solids are wetted prior to disposal	Minefill	Groundwater monitoring around disposal site	None	Ash deposition in absorber due to poor atomization Atomizer drive system failures Control difficulties during load change
Riverside	Lime-Spray Drying	All process equipment	60/40	N/A	About 15-20% moisture added to dry solids prior to disposal	Landfill	Unknown	None	More redundancy and flue gas bypass required
Scholz	Chiyoda Thoroughbred 121	Vacuum filter	0/100	Wet stacking & vacuum filtration	None	Gypsum stack	Groundwater monitoring around disposal site	Continuous	Inability to operate closed loop

5 WASTE CHARACTERIZATION

5.1 Introduction

This chapter presents data on the physical, chemical and radiochemical characteristics of makeup water, wastewaters and sludge samples collected from several FGD systems visited during this study. Also presented are the results of bench scale batch leaching studies on the FGD sludges using a procedure developed by Environment Canada's Wastewater Technology Centre (19).

5.2 Sample Procurement and Analytical Methods

A summary of the wastewater and sludge samples collected at each site is presented in Table 12. Figures 10 through 12 show the locations where grab samples were collected at each of the sites. The following sample volumes were collected:

- 15 L of wastewater for chemical analysis;
- 45 L of wastewater for treatability tests; and
- 18 kg of solid waste for chemical analysis and leaching tests.

TABLE 12 SAMPLE COLLECTION SUMMARY

Generating Station/Location	Date Sampled	FGD System	Samples Collected
Sherburne County Station Northern States Power Co. Becker, Minnesota	October 15/82	Limestone/Alkaline Fly Ash	Thickener overflow Scrubber makeup Recycle water Ash pond overflow Sludge from disposal pond
Milton R. Young Station Minnkota Power Co-operative Inc., Center, North Dakota	October 13/82	Limestone/Alkaline Fly Ash	Blowdown sump Disposal site runoff Pond return water Sludge from disposal site
Coyote Station Montana - Dakota Utilities Co., Beulah, North Dakota	October 14/82	Spray Dry Process/Sodium Carbonate Reagent	Makeup water Disposal site runoff Fly ash/FGD solids from baghouse hoppers Cured FGD solids from disposal site

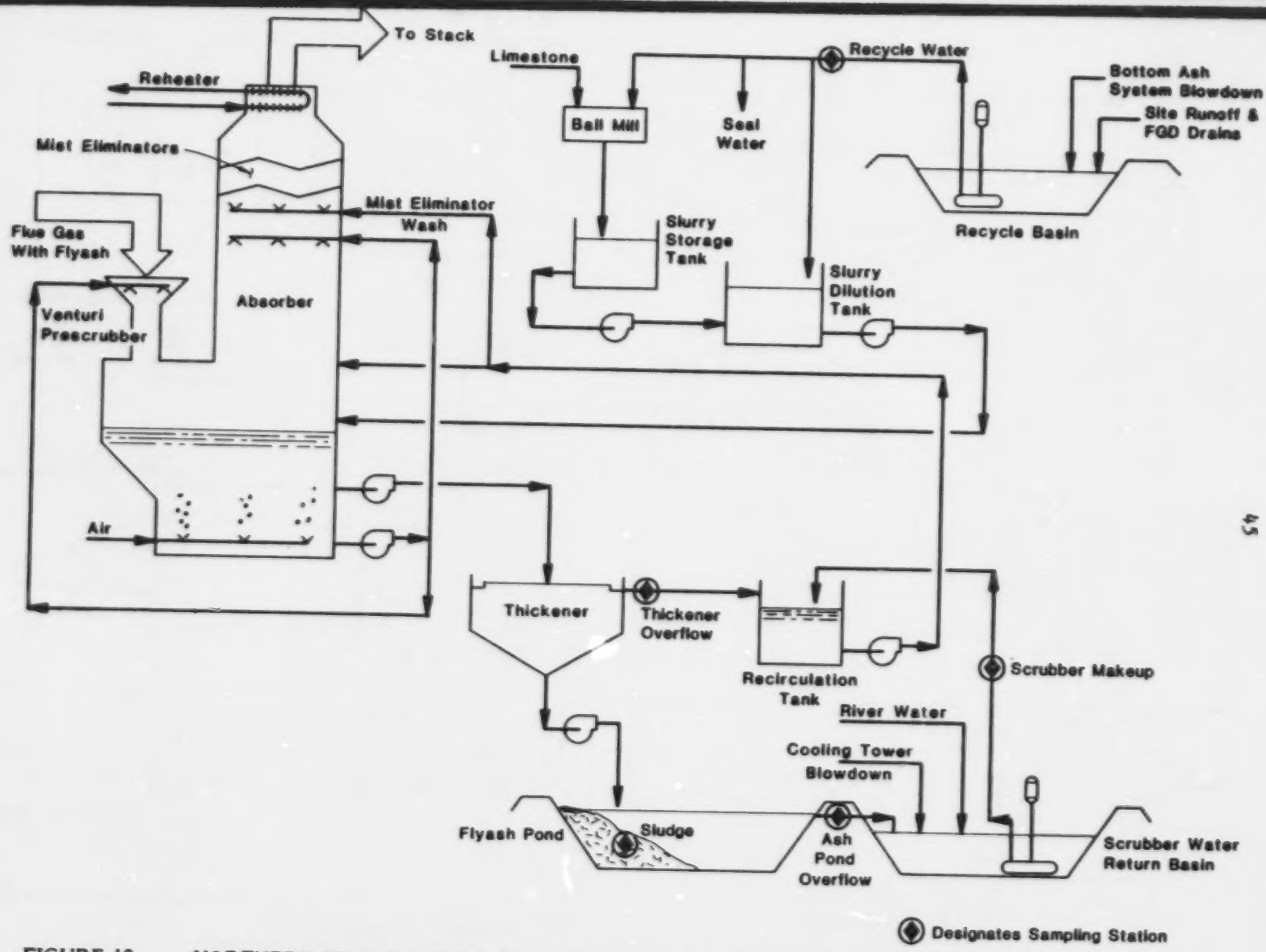


FIGURE 10 NORTHERN STATES POWER CO., SHERBURNE COUNTY GENERATING STATION SAMPLING LOCATIONS

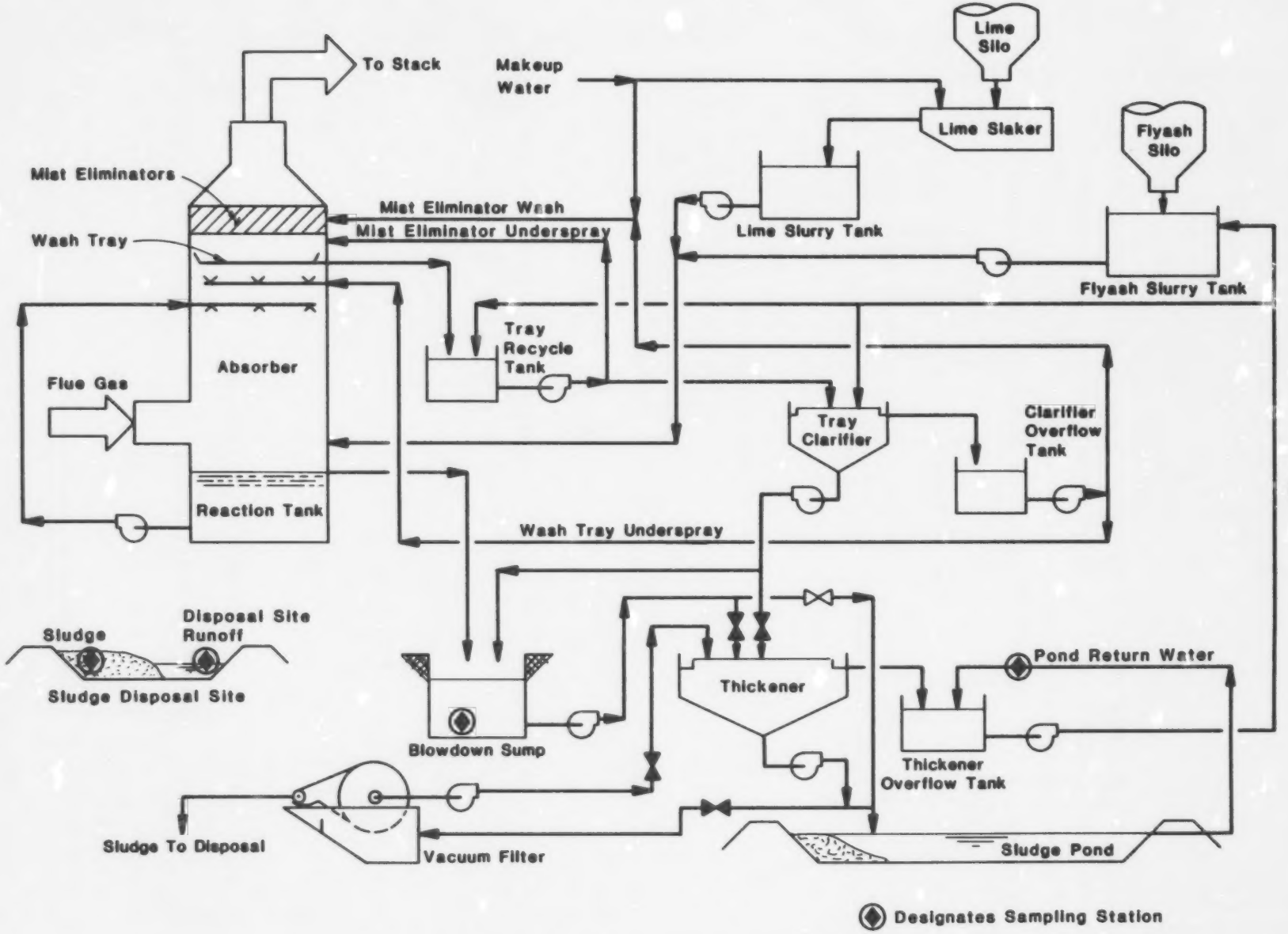


FIGURE 11 MINNKOTA POWER COOPERATIVE INC., MILTON R. YOUNG GENERATING STATION SAMPLING LOCATIONS

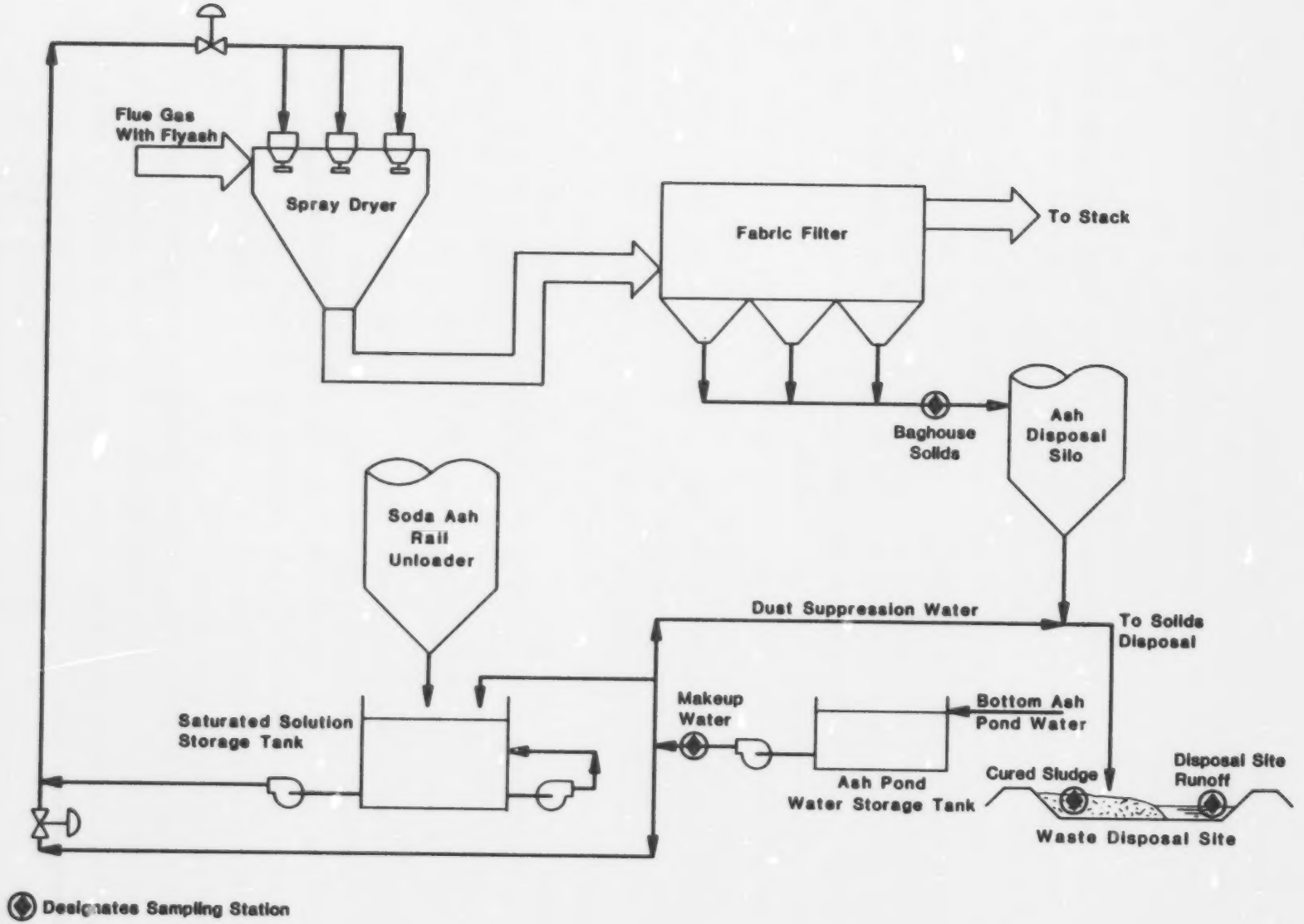


FIGURE 12 MONTANA - DAKOTA UTILITIES CO. COYOTE STATION SAMPLING LOCATIONS

Liquid samples were drawn from runoff containment ponds at solid waste disposal sites. Wastewater samples were not filtered or fixed at the site. All samples were air freighted to laboratories in Ontario and were usually received within 48 hours of shipment.

The laboratories and analytical methods used for the analyses of the samples are outlined in Table 13. The liquid samples were preserved upon arrival at the laboratory, using methods recommended by the U.S. Environmental Protection Agency (20). All samples were refrigerated after shipment and prior to analysis in the laboratory. All analyses were completed on unfiltered samples.

Generally, laboratory analyses for non-metal parameters were performed using procedures outlined in Standard Methods for the Examination of Water and Wastewater (21). Ion chromatography techniques were used to quantify major anion and cation concentrations. The primary method of analysis for total metals was the induced coupled plasma (ICP) scan. Those metals not analyzed using ICP because of unacceptably high detection limits, specifically arsenic, lead, lithium, mercury, selenium, and sodium, were analyzed using atomic absorption (AA) following the appropriate digestion procedure.

The sludge from Milton Young and the cured solids from Coyote Generating Station were collected at abandoned mines which served as solid waste disposal sites. Dry solids from Coyote Generating Station were collected at the baghouse hoppers. The Sherburne Generating Station sludge was sampled at the FGD sludge/fly ash disposal pond. Samples of FGD sludges were refrigerated upon arrival at the laboratory, until leaching tests were conducted. The non-metals in the sludge were analyzed by x-ray diffraction and the metals by the induced coupled plasma scan, atomic absorption, x-ray diffraction, and neutron activation analysis.

5.3 Makeup Water and Wastewaters

5.3.1 Wastewater Characteristics. Results of the analyses of makeup water and wastewater samples from Sherburne, Milton Young, and Coyote stations are provided in Appendix B, Table B-1. At the Sherburne Generating Station, samples collected comprised thickener overflow, scrubber makeup, recycle water and ash pond overflow. The water from the recycle basin which contained discharges from plant drains and bottom ash blowdown was of relatively high quality. It serves both as seal water and dilution water to the FGD system. The thickener overflow consistently had the highest concentrations of all parameters analyzed. The quality of the ash pond overflow was similar to that of the scrubber makeup which consists of ash pond overflow, cooling tower blowdown and river water.

TABLE 13 SAMPLE ANALYSES SUMMARY

Generating Stations	Analyses	Sludge		Wastewater	
		Lab	Method	Lab	Method
Sherburne	Non-Metals	DECS	X	DECS	WC/IC
	Metals	BARR/DECS/SRC	ICP/X/AA/NAA	BARR/DECS/SRC	ICP/AA/NAA
	Radionuclides	SRC/WTC	AS/GS/NAA	SRC/WTC	AS/GS/NAA
Milton R. Young	Non-Metals	DECS	X	DECS	WC/IC
	Metals	BARR/DECS/SRC	ICP/X/AA/NAA	BARR/DECS/SRC	ICP/AA/NAA
	Radionuclides	SRC/WTC	AS/GS/NAA	SRC/WTC	AS/GS/NAA
Coyote	Non-Metals	DECS	X	DECS	WC/IC
	Metals	BARR/DECS/SRC	ICP/X/AA/NAA	BARR/DECS/SRC	ICP/AA/NAA
	Radionuclides	SRC/WTC	AS/GS/NAA	SRC/WTC	AS/GS/NAA

Legend:
DECS - Dearborn Environmental Consulting Services
BARR - Barringer Magenta
SRC - Saskatchewan Research Council
WTC - Wastewater Technology Centre
WC - Wet Chemistry
AA - Atomic Absorption
ICP - Induced Coupled Plasma
IC - Ion Chromatrography
AS - Alpha Spectroscopy
GS - Gamma Spectroscopy
NAA - Neutron Activation Analysis
X - X-Ray Diffraction

Wastewater samples from the Milton Young Generating Station consisted of blowdown sump water, sludge pond return water and sludge runoff/leachate from the disposal site. The sample taken from the blowdown sump was allowed to settle and only the supernatant was analyzed. Thus, the contribution of the suspended solids originally present in the wastewater is not accounted for in the analytical results. The pond return water generally had lower contaminant concentrations than the sludge runoff/leachate. It is possible that water percolating through the abandoned coal mine adds to the dissolved solids in the runoff/leachate from the FGD sludge disposed in the mine.

Samples collected at the Coyote Generating Station included the makeup water and the disposal site runoff/leachate. The makeup water, which originates from the bottom ash system, had elevated levels of certain elements likely due in part to contact with the bottom ash. However, levels of dissolved solids, sulphate, potassium, and sodium were significantly lower than in the sludge runoff.

5.3.2 Comparison of Wastewaters to Proposed Discharge Criteria. In Table 14 analyses of disposal site runoff samples from the Milton Young and Coyote Generating Stations and the ash pond overflow from the Sherburne Generating Station are compared to the pH, metals and suspended solids concentrations specified in the wastewater criteria for discharge proposed by Environment Canada, Environmental Protection Service, in the Steam Electric Environmental Code of Practice (22).

An overall perspective of the data in Table 14 reveals that all three wastewaters have similar characteristics. In general, low concentrations of chromium, copper, iron, nickel and zinc were present. None of these wastewaters were either very acidic or very alkaline. The suspended solids concentrations, as determined by filtration through a 1.2 μ filter paper (21) were four to five times greater than the discharge criterion in all three cases.

In the case of the Sherburne Generating Station overflow, concentrations of all the parameters, except suspended solids, fell within the limits of the discharge criteria. The disposal site runoff from the Coyote Generating Station did not contain any metals in excess of the discharge criteria. However, the pH was slightly alkaline. The suspended solids concentration was about five times greater than the criterion. A comparison of disposal site runoff from the Milton Young Generating Station to the criteria shows that the parameters exceeding the criteria were iron, nickel and suspended solids. Their concentrations were about three, two and five times higher, respectively, than the criteria.

TABLE 14 COMPARISON OF ENVIRONMENT CANADA'S PROPOSED DISCHARGE CRITERIA WITH POTENTIAL WASTEWATERS FROM FGD SYSTEMS

Parameter	Discharge Criteria	Sherburne Ash Pond Overflow	Milton Young Disposal Site Runoff	Coyote Disposal Site Runoff
pH (units)	6.5 - 9.5	8.1	7.3	9.9
Chromium	0.5	0.24	0.24	< 0.01
Copper	0.5	0.024	0.013	< 0.008
Iron	1.0	0.56	2.82	0.28
Nickel	0.5	0.24	0.90	< 0.05
Zinc	0.5	< 0.05	0.23	0.23
Suspended Solids	25	103	126	136

Note: All analyses reported in mg/L except where noted. All metal concentrations are total values, determined by analysis of unfiltered samples.

5.3.3 **Treatability Study.** The disposal site runoff from Milton Young Generating Station was selected for a treatability study because it exceeded the proposed discharge criteria with respect to two metals. The parameters of interest were iron (2.82 mg/L), nickel (0.90 mg/L), and suspended solids (126 mg/L).

An initial attempt at improving the quality of the disposal site runoff involved filtering the sample through a 0.45 micron filter. As well as removing suspended solids, this procedure reduced the iron to an acceptable level of 0.3 mg/L. The nickel, however, being present in soluble form, remained at 0.90 mg/L.

A second treatment approach for the sludge runoff involved coagulation/flocculation using the ASTM jar test procedure (23). Addition of sodium hydroxide (NaOH) to pH values ranging from 7.3 to 9.5 did not reduce the nickel concentration sufficiently to meet the discharge criteria. Coagulation with hydrated lime ($Ca(OH)_2$) was successful; however, the pH had to be raised to 9.5 before the nickel concentration decreased to an acceptable level of 0.17 mg/L.

Characterization of the raw and lime-treated sludge runoff is shown in Table 15. Following treatment, the pH, suspended solids, and five metals in question met the discharge criteria. The concentrations of several other metals such as aluminum, boron, magnesium, manganese, silver, titanium, and vanadium were also reduced as a

result of this treatment. This concentration of calcium, as expected, increased due to the lime treatment. The barium concentration increased as well but remained relatively low. The slight increase in the zinc concentration may represent an analytical error in the technique used.

5.4 FGD Sludges

The physical characteristics of the FGD sludges and waste solids may be observed by examining their photographs, Figures 13 to 16. The brown coloured Milton Young sludge and the grey coloured Sherburne sludge were both soft, wet and pliable. The brownish-grey material from Coyote Generating Station included a powdery sample from the baghouse hoppers, and a cured waste sample from the disposal site, which had a texture similar to dried clay. The constituents present at 0.1 percent or greater in one or more of the sludges are summarized in Table 16.

A detailed comparison is difficult because each station utilized a different coal type and a different FGD system design. Although the chemical characteristics of the FGD solid wastes are variable and site-specific, many of the constituents were present in concentrations of the same order of magnitude in all four sludges.

The Milton Young and Sherburne sludges and Coyote cured material had comparable moisture contents of 29 to 35 percent. The dry material collected from the baghouse hoppers at the Coyote Generating Station had a minimal moisture content of 0.2 percent.

The sulphite-to-sulphate ratio was the lowest (0.002) in the sludge from the Milton Young Generating Station, which uses a lime-alkaline fly ash FGD system, and was also low (0.02) in the sludge from the Sherburne Generating Station which is equipped with a limestone-alkaline fly ash system. The Coyote Generating Station which uses the spray dry/sodium carbonate system produced a dry sludge with a SO_3:SO_4 ratio of 0.12 and a cured sludge with a ratio of 0.15.

In terms of specific constituents, the Milton Young and Sherburne Generating Station sludges contained high levels of sulphate (31 to 33 percent) and calcium (13 to 17 percent), compared to the Coyote material with a sulphate content of 30 percent and calcium of 9 percent. The Coyote Generating Station sludges, however, had much higher sodium and magnesium contents. Iron and aluminum were present in relatively high concentrations in all the sludges, and were highest in the Milton Young sludge. Sludges from all three stations contained comparable levels of silica, strontium, and titanium, averaging 10 percent, 0.32 percent and 0.28 percent, respectively. Most parameters were

TABLE 15 ANALYSIS OF RAW AND LIME-TREATED SLUDGE RUNOFF SAMPLES FROM THE MILTON YOUNG GENERATING STATION

Parameter*	Raw Sample	Treated Sample**	Proposed Discharge Criteria
pH (units)	7.3	9.5	6.5 - 9.5
Suspended Solids	126	3	25
Total Dissolved Solids	45 000	-	
Total Phosphate	1.9	1.7	
Aluminum	3.92	< 0.01	
Barium	< 0.005	0.048	
Beryllium	< 0.0005	< 0.0005	
Boron	346.0	227.0	
Cadmium	0.07	0.04	
Calcium	442.00	504.0	
Chromium	0.24	< 0.01	0.5
Cobalt	0.05	< 0.05	
Copper	0.013	< 0.008	0.5
Iron	2.82	0.23	1.0
Lead	0.05	0.05	
Magnesium	3200	2550	
Manganese	1.00	< 0.01	
Mercury	< 0.001	< 0.001	
Molybdenum	0.4	0.4	
Nickel	0.90	0.17	0.5
Selenium	< 0.001	< 0.001	
Silica	31.9	< 0.05	
Silver	0.053	0.006	
Sodium	6470	5810	
Strontium	18.0	13.0	
Thorium	< 0.01	< 0.01	
Titanium	0.135	< 0.005	
Vanadium	0.102	0.082	
Zinc	0.23	0.28	0.5
Zirconium	< 0.05	< 0.05	

* All analyses in mg/L except pH.
** Coagulation/flocculation with $Ca(OH)_2$.

FIGURE 13 MILTON YOUNG GENERATING STATION FGD SLUDGE

FIGURE 14 SHERBURNE GENERATING STATION FGD SLUDGE

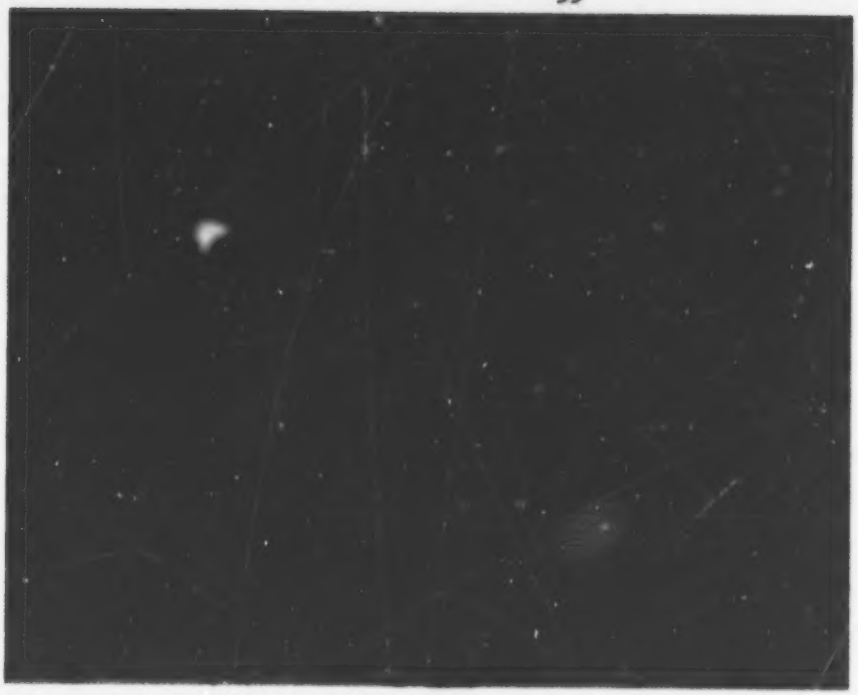


FIGURE 15 COYOTE GENERATING STATION DRY FGD SOLID WASTE

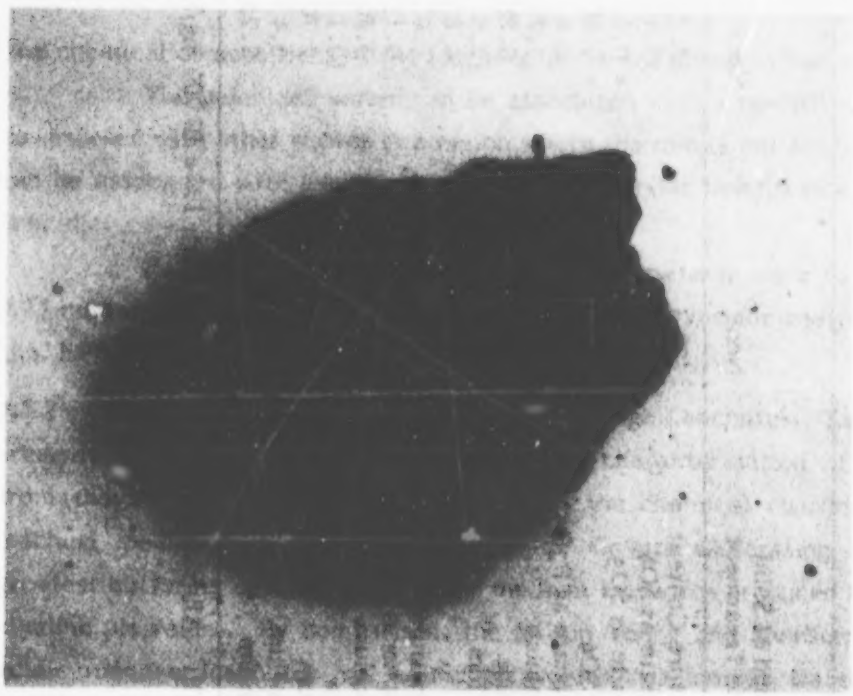


FIGURE 16 COYOTE GENERATING STATION CURED FGD SOLID WASTE

TABLE 16 PHYSICAL PARAMETERS AND PRINCIPAL CONSTITUENTS OF FGD SOLID WASTES FROM SHERBURNE, MILTON YOUNG, AND COYOTE GENERATING STATIONS

Generating Station/ FGD System	Sherburne/ Limestone-Alkaline Fly Ash	Milton Young/ Lime-Alkaline Fly Ash	Coyote/Spray Dry Process - Sodium Carbonate Reagent	
Coal Type	AHV = 20.3 MJ/kg S = 0.8 - 0.9%	AHV = 15 MJ/kg S = 0.4 - 0.9%	AHV = 15.8 MJ/kg S = 0.6 - 1.1%	
Method of Solid Waste Treatment	None	None	Dry Material None	Cured Material Wetted For Dust Suppression
Moisture Content	29.1	35.2	0.2	31.0
Sulphate as SO_4	51.3	55.1	29.0	33.2
Sulphite as SO_3	1.25	< 0.1	3.5	5.1
SO_3/SO_4	0.02	< 0.002	0.12	0.15
Phosphate as P_2O_5	< 0.1	2.4	0.2	2.4
Aluminum	4.8	5.7	5.2	3.8
Barium	0.67	0.88	0.87	0.21
Calcium	16.6	12.6	10.8	8.0
Iron	1.38	6.47	4.06	3.19
Magnesium	0.98	1.63	2.67	2.09
Phosphorus	0.06	0.13	0.11	0.10
Potassium	0.20	0.80	0.60	0.40
Silica	8.1	10.1	10.8	10.5
Sodium	0.60	2.70	11.5	16.0
Strontium	0.21	0.43	0.35	0.27
Titanium	0.24	0.25	0.37	0.27

NOTE: All analyses reported in percent; 1% = 10 000 mg/kg.

present at the lowest concentrations in the Sherburne sludge. There was no appreciable difference in levels of most of the constituents in the cured and uncured Coyote material.

Results of analyses of FGD solid wastes collected in Phase One of this study are presented in Table 17.

5.5 Leaching Tests

As a part of the laboratory program of this study, bench-scale leaching studies were conducted using FGD solid waste samples from Sherburne, Milton Young, and Coyote stations. The leaching protocol used was that developed at Environment Canada's Wastewater Technology Centre (19).

5.5.1 Leaching Procedure. The leaching procedure used to extract samples for analysis from the sludge consisted of four batch extractions using two different leaching media at two liquid to solid ratios. The liquid-solid mixture was rotated for 24 hours at a rate of two revolutions per minute in square plastic bottles so that the sludges were gently tumbled through the leaching media.

The two leaching media used in the procedure were:

- deionized water, and
- acidic buffered solution at a pH of 4.5.

The chemical compositions of the leaching media are shown in Table 18.

The deionized water can be associated with a landfill where the sludge is not co-disposed with other wastes in a region where the rain is not acidic. The acidic solution can be associated with leaching conditions at a similar landfill in a region where the rain is acidic.

Following 24 hours of mixing the supernatants were vacuum-filtered through 0.45 micron filter paper. The filtrate was then preserved for analytical testing following U.S. EPA procedures (20).

5.5.2 Chemical Characteristics of FGD Sludge Leachates. Table B-3 in Appendix B presents the detailed results of the chemical characterization of laboratory leachates from the four FGD sludges. Also shown is the chemical characterization of the two leaching media. The solid waste from the Coyote Generating Station exhibited the greatest buffering capacity; the acidic medium leachates produced had neutral to slightly alkaline pH values. By comparison, the Milton Young and Sherburne Generating Station acidic medium leachates had acidic pH values. Deionized water leachates from the Milton Young waste were neutral, and were slightly alkaline from both Coyote and

TABLE 17 PHYSICAL PARAMETERS AND PRINCIPAL CONSTITUENTS OF FGD SLUDGES FROM CANE RUN, DALLMAN, MARION, AND NEWTON GENERATING STATIONS

Generating Station/ FGD System	Cane Run/ Lime	Dallman/ Limestone Forced Oxidaton	Marion/ Limestone	Newton/ Dual-Alkali
Coal Type	AHV = 25.6 MJ/kg S = 3.75%	AHV = 24.4 MJ/kg S = 3.3%	AHV = 23.3 MJ/kg S = 3.5%	AHV = 27.1 MJ/kg S = 3.0%
Method of Solid Waste Treatment	1:65 Mix of FGD Sludge & Fly Ash; 2% Lime Addition	None	Fly Ash Addition	1:1 Mix of FGD Sludge & Fly Ash; 2% Lime Addition
Moisture Content	36	35	23	30
Sulphite to Sulphate Ratio	5.8	0.37	0.72	1.87
Sulphate as SO_4	6.1	48.5	34.9	18
Sulphite as SO_3	35.6	17.9	25.0	33.6
Chloride	< 0.1	0.1	0.2	0.5
Aluminum	3.96	0.1	3.49	4.06
Calcium	17.9	16.8	15.7	14.4
Iron	4.84	0.15	5.17	5.39
Magnesium	0.39	0.09	1.22	1.21
Potassium	0.76	0.09	0.68	0.75
Silica	7.06	1.03	5.14	5.61
Sodium	0.18	0.06	0.25	0.21
Titanium	0.20	0.007	0.20	0.20

NOTE: All analyses reported in percent; 1% = 10 000 mg/kg.

TABLE 18 CHEMICAL COMPOSITION OF THE LEACHING MEDIA

Leaching Medium	Chemicals Added to Distilled Water	Concentration (M)
Deionized Water	-	-
Acidic Solution	Acetic acid	0.082
	Sodium acetate	0.045

Sherburne wastes. The Coyote Generating Station leachates contained the highest total dissolved solids as well as the highest sulphate and sulphite concentrations. The 4:1 acidic medium leached the most dissolved solids from all the sludges.

The range of concentrations of constituents in the leachates from the three generating stations is shown in Table 19. The wide range in concentrations for several constituents is due to differences in the coals utilized at each station, differences in the reagents used in the FGD processes, the FGD process used, the leaching medium and the liquid-to-solid ratio used during the leaching. Unless negligible, the concentration of the constituent in each medium was subtracted before determining the range. The exceptions were pH and COD. As the acidic medium leachates contained slightly lower COD values than did the medium itself, the upper limit of the range of COD concentrations in Table 19 is presented without subtracting the COD of the medium. It is difficult to ascertain what portion of the COD is imparted by the constituents leached from the sludge. A portion of the COD of the medium may have been absorbed into the sludge during leaching.

Table 20 shows the constituents in the leachates grouped by range of concentration. Only two constituents, cadmium and lead, were not detected in any of the leachates. Sulphate, sulphite, calcium, magnesium and sodium were present at concentrations of 0.1 percent or greater. A similar grouping of constituents leached from solid wastes collected in Phase One of this study is also shown in Table 20.

A more detailed analysis was completed on the leaching of chromium, copper, iron, nickel and zinc under the four test conditions. These metals were selected because discharge criteria have been developed for them by Environment Canada for the Canadian steam electric generating industry. However, other contaminants detected may also be of environmental concern (22).

TABLE 19 RANGE OF CONCENTRATIONS OF CONSTITUENTS IN LABORATORY LEACHATES

Parameter	Unit	Range
pH		4.6 - 12.1
Sp. Conductance	µmhos/cm	2 230 - 47 600*
Total Dissolved Solids	mg/L	2 523 - 77 240
Chemical Oxygen Demand	mg/L	< 10 - 7 450**
"OH" Alkalinity as $CaCO_3$	mg/L	0 - 2 330
Sulphate as SO_4	mg/L	1 560 - 42 400
Sulphite as SO_3	mg/L	< 1 - 12 900
Chloride	mg/L	3.0 - 140*
Fluoride	mg/L	1.4 - 7.0*
Orthophosphate as PO_4^{-3}	mg/L	< 0.05 - 4.5
Aluminum	mg/L	< 0.01 - 137
Arsenic	mg/L	< 0.005 - 5.80
Barium	mg/L	< 0.005 - 0.125
Beryllium	mg/L	<0.0005 - 0.0037
Boron	mg/L	5.86 - 97.8
Cadmium	mg/L	< 0.01
Calcium	mg/L	4.32 - 1 090
Cobalt	mg/L	< 0.05 - 0.10
Chromium	mg/L	< 0.01 - 0.13
Copper	mg/L	< 0.008 - 0.098
Iron	mg/L	< 0.01 - 4.64
Lead	mg/L	< 0.05
Lithium	mg/L	0.1 - 1.4
Magnesium	mg/L	0.04 - 1 081
Manganese	mg/L	< 0.01 - 8.55
Mercury	µg/L	< 1 - 21.0
Molybdenum	mg/L	< 0.3 - 2.3
Nickel	mg/L	< 0.05 - 0.28
Phosphorus	mg/L	< 0.6 - 2.3
Potassium	mg/L	< 1 - 516
Selenium	mg/L	< 0.005 - 3.1
Silica	mg/L	0.28 - 92.4*
Silver	mg/L	0.005 - 0.012
Sodium	mg/L	14 - 22 400*
Strontium	mg/L	4.47 - 41.7
Thorium	µg/L	< 3.5 - 5.7
Titanium	mg/L	< 0.005 - 0.807
Uranium	µg/L	< 3.0 - 95
Vanadium	mg/L	< 0.005 - 1.67
Zinc	mg/L	< 0.05 - 0.38
Zirconium	mg/L	< 0.05 - 0.08

* Concentration of constituent in blank (medium) was subtracted before determining the range.

** Concentration of COD in blank not subtracted.

TABLE 20 CONSTITUENTS IN LABORATORY LEACHATES GROUPED BY RANGE OF CONCENTRATIONS

Group	Phase One	Phase Two
I: Constituents present at concentrations less than detection limit	Silver, Zirconium	Cadmiun, Lead
II: Constituents present at concentrations less than 1 mg/L but greater than detection limit	Beryllium, Cadmium, Cobalt, Chromium, Copper, Lead, Mercury Nickel, Selenium, Titanium, Vanadium	Barium, Beryllium, Cobalt, Chromium, Copper, Mercury, Nickel, Silver, Thorium, Titanium, Uranium, Zinc, Zirconium
III: Constituents present at concentrations less than 10 mg/L but greater than 1 mg/L	Arsenic, Barium, Fluoride, Orthophosphate, Lithium, Manganese, Molybdenum, Phosphorus, Strontium, Zinc	Arsenic, Iron, Fluoride, Orthophosphate, Lithium, Manganese, Molybdenum, Phosphorus, Selenium, Vanadium
IV: Constituents present at concentrations less than 100 mg/L but greater than 10 mg/L	Aluminum, Boron, Iron Silica	Boron, Silica, Strontium
V: Constituents present at concentrations less than 1000 mg/L but greater than 100 mg/L	Sulphite, Chloride, Magnesium, Potassium	Chloride, Aluminum, Potassium
VI: Constituents present at concentrations greater than 1000 mg/L	Sulphate, Calcium, Sodium	Sulphate, Sulphite, Calcium, Magnesium, Sodium

Tables 21, 22, 23 and 24 present the concentrations, the releases and the leaching efficiencies of each metal under each of the leaching conditions. The concentration is the weight of a chemical specie in the leachate per volume of leachate. The release is the weight of a chemical specie in the leachate per weight of the sludge sample. The leaching efficiency is the weight of a chemical specie in the leachate per weight of a chemical specie in the sludge sample. The leaching efficiency is highly dependent on the analytical technique as, in its calculation, the weight of a chemical specie in the sludge is required. This is often a difficult analytical procedure and is dependent on the digestion procedure used.

TABLE 21 BEHAVIOUR OF FIVE METALS IN LEACHATES PRODUCED FROM MILTON YOUNG GENERATING STATION FDG SLUDGE

Parameter		Deionized Medium		Acidic Medium	
		4:1	20:1	4:1	20:1
Chromium	Concentration (mg/L)	<0.01	<0.01	0.02	<0.01
	Release (mg/g)	<0.04	<0.20	0.08	<0.20
	Leaching Efficiency (%)	-	-	0.21	-
Copper	Concentration (mg/L)	0.011	0.012	0.020	0.098
	Release (mg/g)	0.044	0.240	0.080	1.96
	Leaching Efficiency (%)	0.05	0.30	0.10	2.44
Iron	Concentration (mg/L)	0.09	0.06	0.25	4.64
	Release (mg/g)	0.36	1.2	1.0	92.8
	Leaching Efficiency (%)	0.001	0.003	0.002	0.22
Nickel	Concentration (mg/L)	0.09	0.09	0.28	0.14
	Releast (mg/g)	0.36	1.8	1.1	2.8
	Leaching Efficiency (%)	0.09	0.44	0.27	0.69
Zinc	Concentration (mg/L)	0.10	0.08	0.38	0.26
	Release (mg/g)	0.40	1.6	1.5	5.2
	Leaching Efficiency (%)	0.36	1.5	1.4	4.7

Note: Concentration = weight of chemical specie in leachate per volume of leachate.
Release = weight of chemical specie in leachate per weight of total sludge sample.
Efficiency = weight of chemical specie in leachate per weight of chemical specie in sludge sample.

TABLE 22 BEHAVIOUR OF FIVE METALS IN LEACHATES PRODUCED FROM COYOTE GENERATING STATION DRY (BAGHOUSE) FGD SOLIDS

Parameter		Deionized Medium		Acidic Medium	
		4:1	20:1	4:1	20:1
Chromium	Concentration (mg/L)	0.06	0.05	<0.01	<0.01
	Release (mg/g)	0.24	1.0	<0.04	<0.20
	Leaching Efficiency (%)	0.48	2.0	-	-
Copper	Concentration (mg/L)	<0.008	<0.008	<0.008	<0.008
	Release (mg/g)	<0.032	<0.160	<0.032	<0.160
	Leaching Efficiency (%)	-	-	-	-
Iron	Concentration (mg/L)	<0.01	0.04	<0.01	0.10
	Release (mg/g)	<0.04	0.80	<0.04	2.0
	Leaching Efficiency (%)	-	0.002	-	0.005
Nickel	Concentration (mg/L)	<0.05	<0.05	<0.05	0.13
	Releast (mg/g)	<0.20	<1.0	<0.20	2.6
	Leaching Efficiency (%)	-	-	-	8.7
Zinc	Concentration (mg/L)	<0.05	<0.05	<0.05	0.08
	Release (mg/g)	<0.20	<1.0	<0.20	1.6
	Leaching Efficiency (%)	-	-	-	1.5

Note: Concentration = weight of chemical specie in leachate per volume of leachate.
Release = weight of chemical specie in leachate per weight of total sludge sample.
Efficiency = weight of chemical specie in leachate per weight of chemical specie in sludge sample.

TABLE 23 BEHAVIOUR OF FIVE METALS IN LEACHATES PRODUCED FROM COYOTE GENERATING STATION CURED FGD WASTE

Parameter		Deionized Medium		Acidic Medium	
		4:1	20:1	4:1	20:1
Chromium	Concentration (mg/L)	0.20	0.01	<0.01	<0.01
	Release (mg/g)	0.08	0.20	<0.04	<0.20
	Leaching Efficiency (%)	0.39	0.97	-	-
Copper	Concentration (mg/L)	<0.008	0.008	<0.008	<0.008
	Release (mg/g)	<0.032	0.160	<0.032	<0.160
	Leaching Efficiency (%)	-	-	-	-
Iron	Concentration (mg/L)	<0.01	<0.01	<0.01	0.02
	Release (mg/g)	<0.04	<0.20	<0.04	0.40
	Leaching Efficiency (%)	-	-	-	0.002
Nickel	Concentration (mg/L)	<0.05	<0.05	<0.05	0.07
	Release (mg/g)	<0.20	<1.0	<0.20	1.4
	Leaching Efficiency (%)	-	-	-	6.8
Zinc	Concentration (mg/L)	<0.05	<0.05	<0.05	0.07
	Release (mg/g)	<0.20	<1.0	<0.20	1.4
	Leaching Efficiency (%)	-	-	-	2.3

NOTE: Concentration = weight of chemical specie in leachate per volume of leachate.
Release = weight of chemical specie in leachate per weight of total sludge sample.
Efficiency = weight of chemical specie in leachate per weight of chemical specie in sludge sample.

TABLE 24 BEHAVIOUR OF FIVE METALS IN LEACHATES PRODUCED FROM SHERBURNE GENERATING STATION FGD SLUDGE

Parameter		Deionized Medium		Acidic Medium	
		4:1	20:1	4:1	20:1
Chromium	Concentration (mg/L)	0.13	0.03	0.05	0.07
	Release (mg/g)	0.52	0.60	0.20	1.4
	Leaching Efficiency (%)	3.7	4.2	1.4	9.9
Copper	Concentration (mg/L)	<0.008	0.010	0.011	0.082
	Release (mg/g)	<0.032	0.200	0.044	1.64
	Leaching Efficiency (%)	-	0.88	0.19	7.23
Iron	Concentration (mg/L)	0.04	0.02	0.11	3.61
	Release (mg/g)	0.16	0.40	0.44	72.2
	Leaching Efficiency (%)	0.002	0.004	0.004	0.74
Nickel	Concentration (mg/L)	0.07	0.07	0.16	0.15
	Release (mg/g)	0.28	1.4	0.64	3.0
	Leaching Efficiency (%)	2.0	9.9	4.5	21
Zinc	Concentration (mg/L)	<0.05	0.05	0.23	0.20
	Release (mg/g)	<0.20	0.20	0.92	4.0
	Leaching Efficiency (%)	-	4.7	4.3	19

NOTE: Concentration = weight of chemical specie in leachate per volume of leachate.
Release = weight of chemical specie in leachate per weight of total sludge sample.
Efficiency = weight of chemical specie in leachate per weight of chemical specie in sludge sample.

The releases and leaching efficiencies demonstrate several trends. Increasing the liquid-to-solid ratio from 4:1 to 20:1 resulted in a higher contaminant release and leaching efficiency because the amount of medium contacting the waste was increased. This phenomenon was less pronounced with the deionized medium leachates.

The acidic medium leachates produced greater contaminant releases and leaching efficiencies, except for chromium under some conditions. Leachate pH, a function of the alkalinity of the sludge and the amount of acid available in the leaching medium to react with the waste, is known to have a major influence on contaminant solubility.

The overall leaching efficiencies shown in Tables 21 to 24 range from 0.001 percent for iron in the 4:1 deionized medium leachate from the Milton Young sludge, to 21 percent for nickel in the 20:1 acidic medium leachate from the Sherburne sludge. For the most part, nickel and zinc were leached more efficiently than chromium, copper and iron.

Little difference was observed between the contaminant releases and leaching efficiencies for the cured and uncured Coyote FGD solid waste. This was further substantiated by the physical behaviour of the cured material. Upon addition of a leaching medium, the cured material readily broke apart and became virtually indistinguishable from the uncured solids. However, a direct comparison of the chemical constituents in the leachates from these two sludges may not be appropriate. At the time the cured material was produced, originally as baghouse solids, the station or FGD system operating conditions may have been slightly different (e.g., different coal seam). Furthermore, since the cured material was collected from an abandoned coal mine, some contaminants that leached out of the surrounding spoils and tailings may have been incorporated into the material, while exposure to climatic factors, such as precipitation, may have already leached out some portion of the constituents originally present.

Examination of Tables 14 and 19 allows comparison of the metal concentrations in wastewaters collected at the actual disposal sites to those produced in the laboratory using solid wastes from the sites. The concentrations of the metals in the actual leachates fell within, or in a few cases, slightly above the concentration range for the laboratory leachates. In addition, the chromium, copper, iron, nickel and zinc concentrations in the laboratory leachates shown in Tables 21 to 24 were below Environment Canada's proposed discharge criteria, with the exception of iron concentrations in the Milton Young and Sherburne Generating Stations' 20:1 acidic medium leachates.

5.6 Radiochemical Characteristics

Radiological analyses were performed on all FGD system wastewaters, sludges and laboratory leachates. The concentrations of uranium and thorium, as determined by neutron activation analysis are presented in Appendix B, Tables B-1, B-2 and B-3. Gross alpha and beta determinations were performed on the Milton Young and Coyote laboratory leachates, and on solid wastes and wastewaters from all three stations, shown in Appendix B, Table B-4(a). The activities of certain uranium and thorium decay series radionuclides were determined by alpha and gamma spectroscopy for the Milton Young wastewaters and leachates, and are shown in Appendix B, Table B-4 (b).

The solid wastes contained 8 to 12 μg/g of thorium and 5 to 9 μg/g of uranium, corresponding to specific activities of 33 to 49 Bq/g and 63 to 113 Bq/g, respectively. These values are comparable to those found in coal fly ash (24). The uranium concentration ranged from less than 3.0 to 53 μg/L in the wastewaters, and less than 3.0 to 95 μg/L in the laboratory leachates. These concentrations, converted to specific activity, yield less than 38 to 667 Bq/L and less than 38 to 1195 Bq/L, respectively. The thorium concentrations in the wastewaters and leachates were generally undetectable. This could be explained by the fact that thorium becomes soluble only at acidic pH values.

Gross activities of the solid wastes (Appendix B, Table B-4) ranged from 0.37 Bq/g for both alpha and beta emissions for the Sherburne sludge, to 1.5 Bq/g, alpha and 1 Bq/g, beta, for the Milton Young sludge. Gross alpha activity was undetectable in the liquid portion of all the wastewaters, and ranged from less than 0.04 to 0.26 Bq/L in the filtered suspended solids portion of the wastewaters. The gross beta activity, on the other hand, was higher in the liquid portion of the wastewaters than in the solid portion. The wastewaters with relatively higher total beta activities measured on unfiltered samples were the Milton Young blowdown sump water (21 Bq/L), Milton Young disposal site runoff (0.37 Bq/L), and Coyote disposal site runoff (5.5 Bq/L).

5.7 Quality Control

To verify the reproducibility of the results, certain tests and analyses were repeated. The results are presented in Table B-5 of Appendix B. The leaching protocol was performed on all four sludges a second time under selected leaching conditions. The analyses of these leachates were compared to the corresponding analyses from the previous testing.

The majority of analyses were found to be highly reproducible. The few times that the repeated analyses deviated (by 25 to 50 percent), the concentration was usually close to the detection limit. The protocol and analytical methods can be considered reproducible when possible errors due to sampling method, leaching, transportation, and analyses are taken into account.

The analyses repeated on the dry sludge sample from the Coyote Generating Station substantiated the reproducibility of the analytical method used for solid samples. Further, the results confirmed the homogeneity of the sludge sample collected.

5.8 Discussion and Conclusions

Based on the results of the physical, chemical and radiochemical characterization studies of the FGD system sludges wastewaters and laboratory leachates of Phase One and Phase Two of this study, several conclusions can be made:

- FGD sludges from the eastern U.S. generating stations examined in Phase One of this study, which burn medium to high sulphur bituminous coals, had distinctly higher sulphite concentrations and SO_3:SO_4 ratios than the sludges from the western U.S. generating stations examined in Phase Two, which burn low sulphur subbituminous and lignitic coal.
- The FGD sludges from Phase Two contained higher concentrations of most metals. This may be due to differences in the amount and composition of fly ash contained in the sludges.
- The sludge disposal area runoff and overflow samples collected in Phase Two had pH values within the proposed discharge criteria range of 6.5 to 9.5, except for the Sherburne ash pond overflow which had a slightly alkaline pH of 9.9. The Phase One ash disposal area overflows generally had highly acidic or highly alkaline pH values.
- The three ash disposal area wastewaters examined in Phase Two required some form of suspended solids removal. Of the three wastewaters, only the Milton Young sludge runoff contained primary metal concentrations higher than the proposed discharge criteria. Most of the iron was present in a suspended rather than dissolved form and was removed by filtration. The removal of nickel to an acceptable level, however, required coagulation/flocculation with hydrated lime at a pH of 9.5.
- Chromium, copper, nickel and zinc were present in the laboratory leachates at concentrations of less than 1 mg/L and iron was present at concentrations between 1 and 10 mg/L. Constituents present at concentrations greater than 1000 mg/L included sulphate, sulphite, calcium, magnesium and sodium.

- The overall leaching efficiencies ranged from 0.001 percent to 21 percent. The acidic medium and higher liquid-to-solid ratio generally produced higher leaching efficiencies. There was no significant difference in the contaminant release or leaching efficiencies of the cured versus the uncured Coyote solid waste leachates.
- The relative leachability of primary metals in the Phase Two study generally had nickel leached most efficiently, followed by zinc, chromium, copper and iron; in Phase One, copper was leached most efficiently, followed by nickel, zinc, chromium, and iron.

6 FGD PROCESS CONSIDERATIONS

6.1 Introduction

This chapter comprises three sections. In the first, material balances are presented for FGD systems considered applicable with each of the Canadian coal types considered in Phases One and Two of this study. The water requirements and quantities of wastewater and solid waste generated by each process are identified and compared for each coal type. In the second section, several FGD process design alternatives are discussed, particularly those which may affect the quantity or characteristics of waste produced or overall system performance. Finally, FGD system reliability is addressed. Factors that have contributed to historically low reliabilities are identified and measures that can be taken to improve performance are described.

6.2 Material Balances

6.2.1 Basis for Material Balances. Material balances were developed for different FGD systems applied to a hypothetical 500-MW(e), coal-fuelled station operating at 37 percent energy conversion efficiency. Four coal types considered representative of different Canadian regions were used in the analysis. The characteristics of these fuels are summarized in Table 25. While actual characteristics will depend on mine-specific conditions, the proximate analyses assumed for this study are adequate to illustrate the effects of coal composition of FGD system selection and material balances. In Phase One of this study, an Ontario coal-fuelled station was used as the basis for the evaluation, while in Phase Two, eastern high and medium sulphur coals and a western lignite case were examined.

It was assumed that 92 percent of the sulphur in the coal would evolve as SO_2 in the flue gas and that the FGD systems would achieve SO_2 removal efficiencies corresponding to the National Guidelines for Atmospheric Emissions from New Thermal Power Generation Facilities (i.e., 258 ng/J or 90 percent removal, whichever is least stringent). SO_2 removal requirements are shown as a function of coal sulphur content and average heating value (AHV) in Figure 17, based on the above criteria.

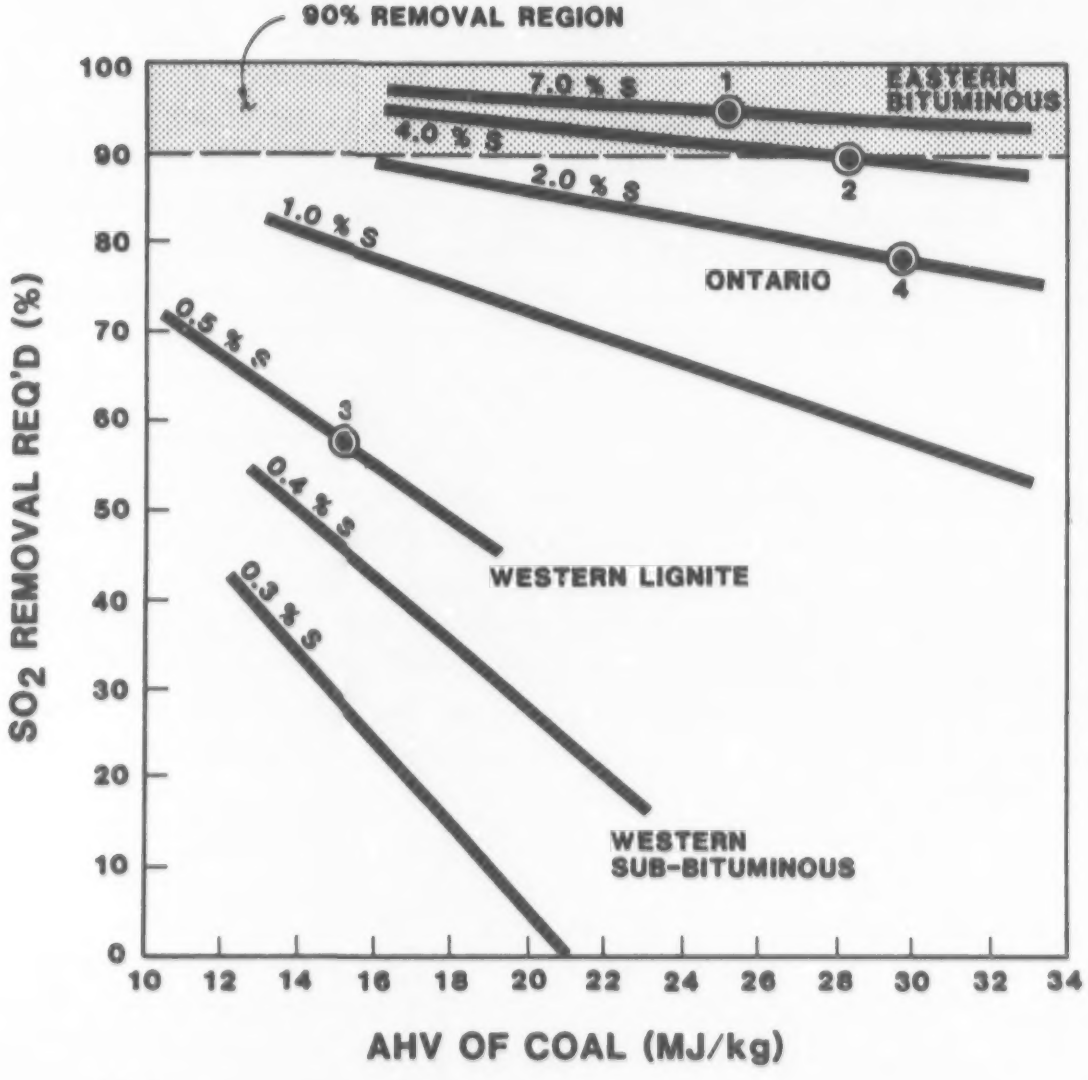
Material balances were not developed for a western subbituminous coal, due to the predicted low SO_2 removals that would be required to meet the emission guidelines. Lower grade western subbituminous coals (i.e., higher sulphur content and lower AHV) would approach the western lignite case assumed in this study.

TABLE 25 SUMMARY OF COAL TYPES AND FGD SYSTEMS EVALUATED

Coal Type*	Typical of	Coal Proximate Analysis S (%)	Ash (%)	Moisture (%)	AHV (MG/kg)	FGD Processes	SO_2 Removal Required** (%)
1)	Eastern High Sulphur Bituminous (New Brunswick)	7.0	20	5	25	- Limestone - Wellman-Lord	90
2)	Eastern Medium Sulphur Bituminous (Nova Scotia)	4.0	12	7	28	- Limestone - Seawater Scrubbing	90
3)	Western Lignite (Saskatchewan)	0.5	13	25	15	- Limestone - Spray Dryer - Alkaline Fly Ash	60
4)	Ontario	2.0	8.5	7	29.6	- Limestone - Lime - Dual Alkali - Wellman-Lord	80

* Western subbituminous coal not evaluated.

** SO_2 removal required to meet National Guidelines for Atmospheric Emissions from New Thermal Power Generation Facilities.



NOTES:

1. 92% of S in coal is assumed to evolve as SO2
2. SO_2 removal requirements are based on National Guidelines for Atmospheric Emissions from Thermal Power Generation Facilities (258 ng/J or 90% removal, whichever is least stringent)
3. ● Denotes coal types considered in this study

FIGURE 17 FLUE GAS SO_2 REMOVAL REQUIREMENTS AS A FUNCTION OF HEAT VALUE AND SULPHUR CONTENT OF COAL

The limestone FGD system was considered a base case for all four coal types as it still comprises the great majority of existing and proposed scrubbers. As the capital and operating costs for the Wellman-Lord process are generally substantially higher than reported for most other processes, it would be primarily applicable to the eastern high sulphur coal case which would produce the greatest quantity of by-product and thus offer the greatest operating costs credit. However, for purposes of comparison to other processes, it was also evaluated for the Ontario station. Seawater scrubbing was only evaluated for the eastern, medium sulphur coal as such a facility would obviously have to be located on the seaboard, and this represents the upper limit of sulphur content to which the process could be applied. Alkaline fly ash and spray drying processes were both considered for the western lignite case. The former is a demonstrated process which has been used on similar coals in the U.S.A., while the latter is a relatively new process that has gained considerable interest in recent years, particularly for the scrubbing of low sulphur coals. Lime and dual alkali processes were also considered in Phase One of this study for the Ontario coal case.

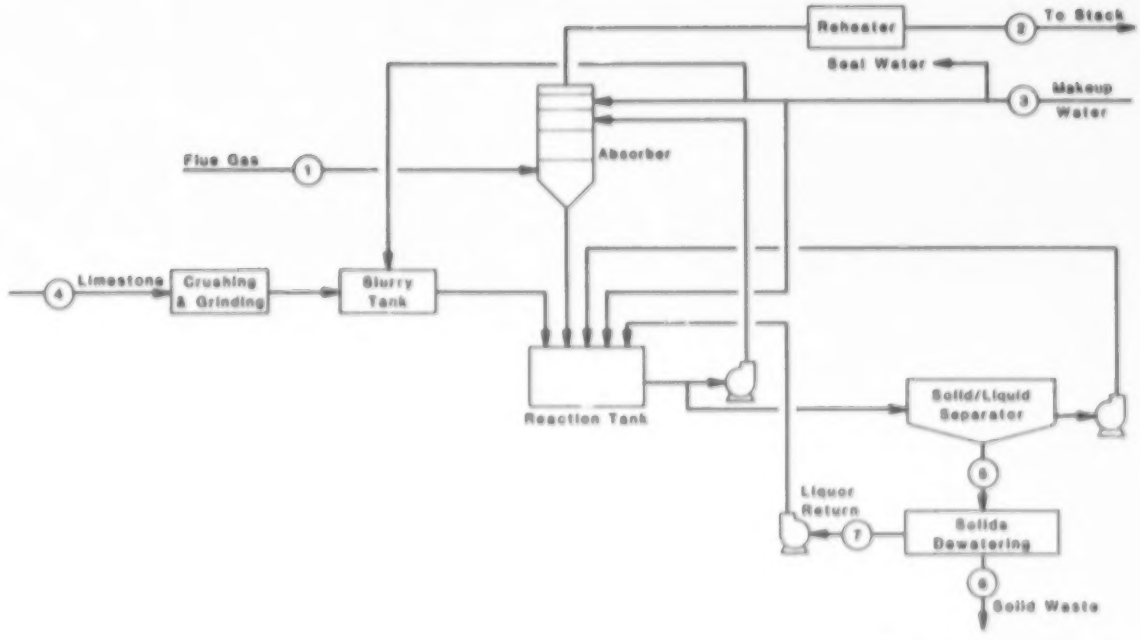
Table 25 summarizes the characteristics of the four coal types examined in this study, their associated SO_2 removal requirements, and the FGD systems considered for each. Several assumptions were made regarding the process design criteria for each FGD system considered. These are summarized in Table 26 and the resulting material balances are presented in Figures 18 through 24.

6.2.2 Reagent Requirements. Figures 18 through 24 illustrate that a limestone FGD system would consume significantly more reagent than any of the other processes considered. The alkaline fly ash and spray dryer processes are both able to take advantage of the natural alkalinity present in the western lignite fly ash. This has a profound effect upon reagent requirements. If it is assumed that the western lignite contains 13 percent available CaO in the ash, and that 75 percent of the ash evolves as fly ash and is captured and fed to the scrubber, the FGD system would operate at a stoichiometric $Ca:SO_2$ ratio of 2.48 for 60 percent SO_2 removal, without any supplementary addition of reagent under normal operating conditions. While similar results are anticipated with the spray dryer process if baghouse and dryer solids are recirculated, very limited data have been reported in the literature regarding operation with solids recycle. A conservative approach was taken and it was assumed that makeup lime would be required at a stoichiometric ratio of 0.5, and solids recirculation would provide the remaining alkalinity necessary to achieve 60 percent SO_2 removal.

TABLE 26 PROCESS DESIGN CRITERIA USED FOR FGD SYSTEM MATERIAL BALANCES

Design Criteria	Lime	Limestone	Dual Alkali	Alkaline Fly Ash	Spray Dryer	Seawater Scrubbing	Wellman-Lord
Reagent Composition	95% CaO 5% Inerts	95% $CaCO_3$ 5% Inerts	Soda Ash-100% Na_2CO_3 Lime-95% CaO	13% available CaO in western lignite	95% CaO 5% Inerts	95% CaO 5% Inerts	Soda Ash-100% $NaCO_3$
Stoichiometric Ratio[a]	1.05	1.15	Na-0.05 Ca-1.05	2.48[b]	0.5[c]	0.8	0.10
Absorbent Feed Concentration	35% $Ca(OH)_2$ in slaker	35% $CaCO_3$ in ball mill	Soda Ash-20% solution Lime-35% $Ca(OH)_2$ in slaker	N/A	35% CaO in slaker	35% in slaker	20% Soda Ash solution
Inlet Flue Gas Temp. (°C)	149	149	149	149	149	149	149
Outlet Flue Gas Temp.	Adiabatic Sat.	Adiabatic Sat.	Adiabatic Sat.	Adiabatic Sat.	10°C Approach to Adiabatic Sat.	25°C	Adiabatic Sat.
% Oxidation from SO_3 to SO_4	15% without forced oxidation, 85% with forced oxidation	15% without forced oxidation, 85% with forced oxidation	15%	100%	35%	100%	10%
Hydrated Solids in Thickened Sludge (wt %)	30%	30%	30%	30%	N/A	N/A	N/A
% Hydrated Solids in Dewatered Waste for Disposal	60% without forced oxidation, 80% with forced oxidation	60% without forced oxidation, 80% with forced oxidation	60%	65%	85%	N/A	60% from centrifuge 100% from dryer
Other	1) 100% of $CaSO_3$ occurs as hemihydrate (i.e., $CaSO_3 \cdot 1/2H_2O$). 2) 100% of $CaSO_4$ occurs as dihydrate (i.e., $CaSO_4 \cdot 2H_2O$). 3) Excess Ca is recarbonated and precipitated as $CaCO_3$.	1) 100% of $CaSO_3$ occurs as hemihydrate (i.e., $CaSO_3 \cdot 1/2H_2O$). 2) 100% of $CaSO_4$ occurs as dihydrate (i.e., $CaSO_4 \cdot 2H_2O$).	1) 100% of $CaSO_3$ occurs as hemihydrate (i.e., $CaSO_3 \cdot 1/2H_2O$). 2) 100% of $CaSO_4$ occurs as dihydrate (i.e., $CaSO_4 \cdot 2H_2O$). 3) Excess Ca is recarbonated and precipitated as $CaCO_3$.	1) 100% of $CaSO_4$ occurs as dihydrate (i.e., $CaSO_4 \cdot 2H_2O$) 2) Fly ash contains 13% available CaO. 3) 75% of coal ash evolves as fly ash.	1) Baghouse & dryer solids are recycled. 2) Fly ash contains 13% available CaO.	1) Gas outlet temp. = seawater inlet to scrubber. 2) Stoichiometric ratio based on vendor-estimated requirement to restore pH to 7.0.	1) 75% of coal ash evolves as fly ash. 2) 99% of fly ash removed by ESP. 3) 100% of residual fly ash is removed in prescrubber. 4) Prescrubber blowdown based on maintaining a 5% solids concentration. 5) Product SO_2 stream contains 10% H_2O. 6) Purge solids contain 80% Na_2SO_4 and 20% Na_2SO_3.

a Stoichiometric Ratio = moles of reagent added/moles of SO_2 removed.
b Ca/S stoichiometric ratio = 2.48 based on available CaO in fly ash, without supplementary reagent addition
c Assuming recycle of baghouse and dryer solids, stoichiometric ratio of 0.5 for reagent lime is assumed for 60% SO_2 removal.

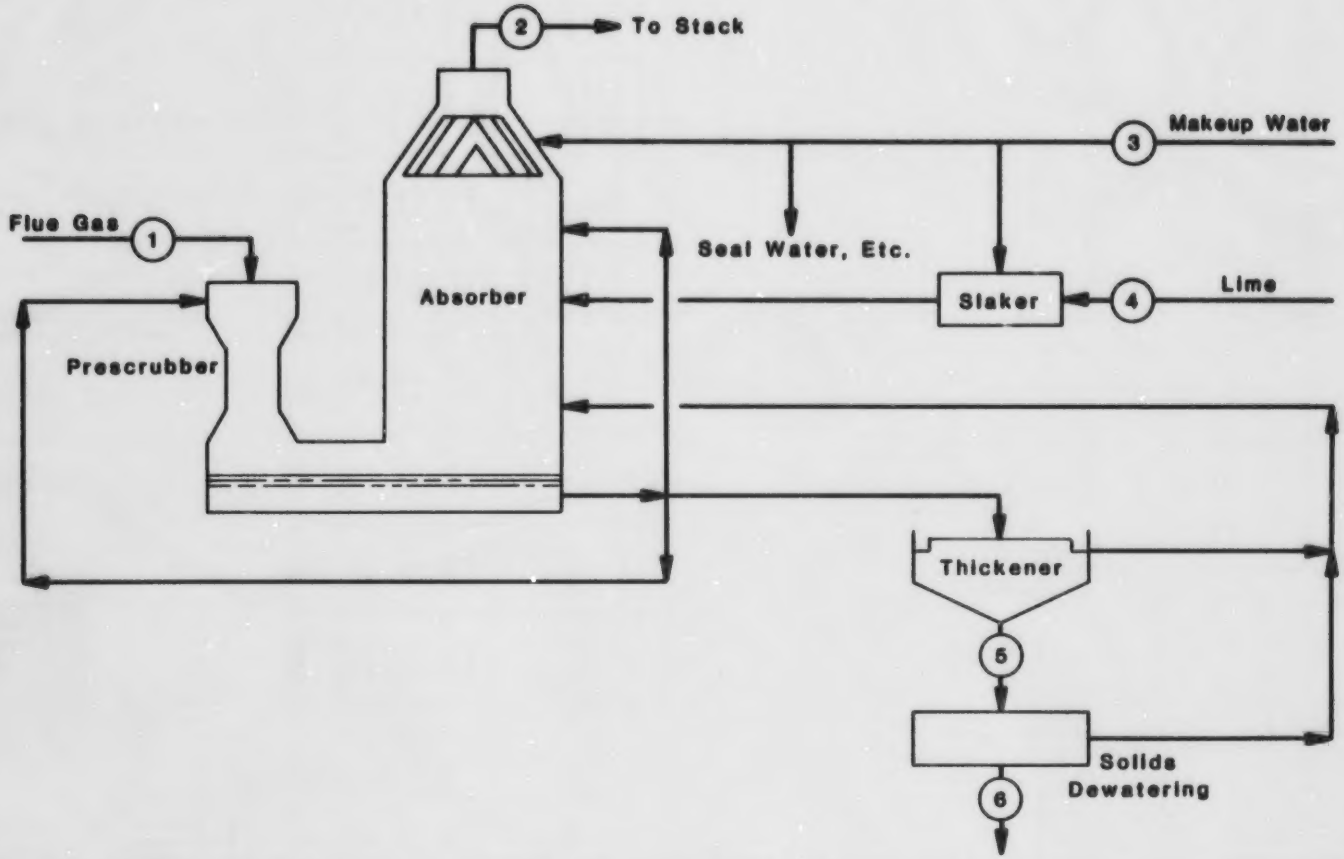


Coal Type	STREAM											
	1			2		3	4	5		6		7
	Gas Flow (Nm^3/min)	Sulphur (kg/min)	Water (kg/min)	Sulphur (kg/min)	Water (kg/min)	Water (kg/min)	Limestone (95% $CaCO_3$) (kg/min)	Solids (kg/min as inerts, $CaCO_3$ & hydrated $CaSO_3$)	Water (kg/min including H_2O of hydration)	Solids (kg/min as inerts, $CaCO_3$ & hydrated $CaSO_3$)	Water (kg/min including H_2O of hydration)	Water (kg/min)
#1												
W/O F.O.	27 630	208.8	1801	20.9	3667	2553	710	931	2248	931	667	1661
W/ F.O.	27 630	208.8	1801	20.9	3667	2320	710	1107	2770	1107	494	2306
#2												
W/O F.O.	22 767	99.3	1855	9.9	3073	1848	336	441	1064	441	330	734
W/ F.O.	22 767	99.3	1855	9.9	3073	1736	336	524	1312	524	220	1092
#3												
W/O F.O.	30 966	24.8	3069	9.9	5037	2026	60	75	180	75	55	130
W/ F.O.	30 966	24.8	3069	9.9	5037	2007	60	93	232	93	39	193
#4												
W/O F.O.	25 250	62	1730	10	3262	1666	168	207	501	207	166	346
W/ F.O.	25 250	62	1730	10	3262	1646	168	247	617	247	103	514

NOTE:
W/O F.O.- Without Forced Oxidation
W/ F.O. - With Forced Oxidation

Coal Type	%S	AHV (MJ/kg)
1. High S Bituminous	7.0	26
2. Med. S Bituminous	4.0	28
3. Western Lignite	0.5	18
4. Ontario	2.0	29.6

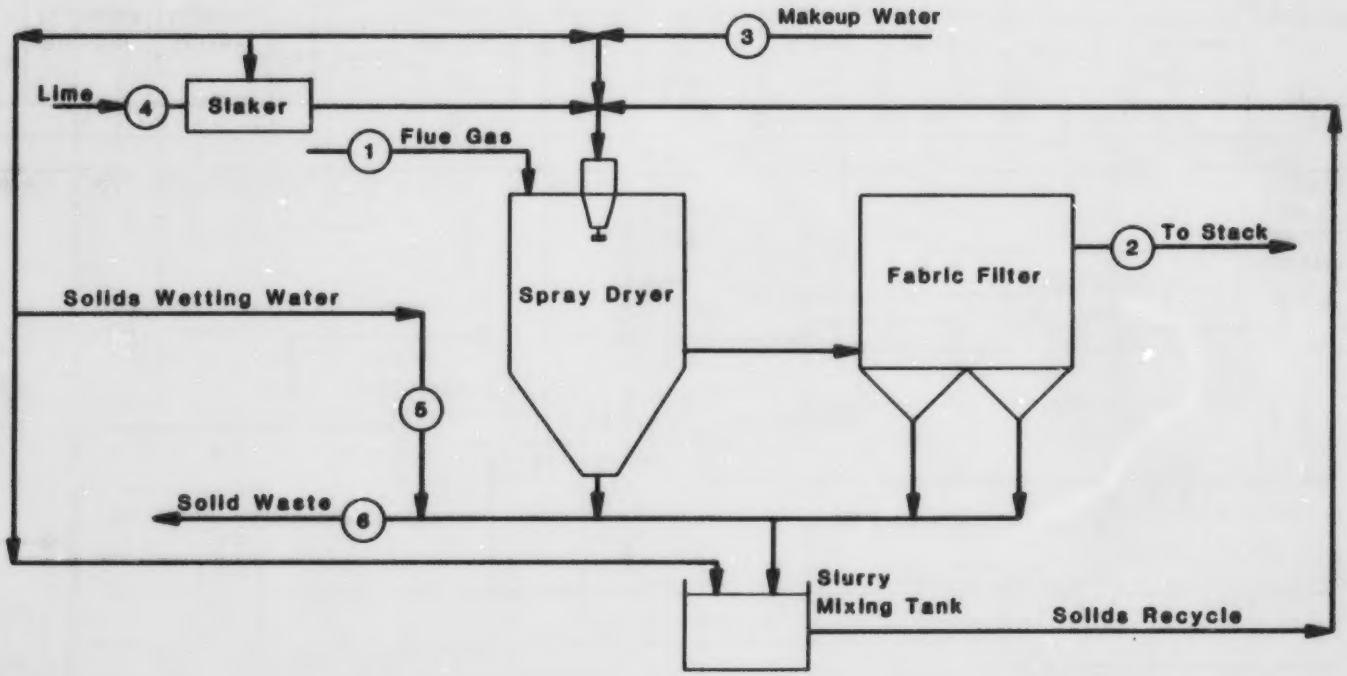
FIGURE 18 MATERIAL BALANCE FOR LIMESTONE SCRUBBING PROCESS



Coal Type	STREAM										
	1			2		3	4	5		6	
	Gas Flow (Nm^3/min)	Sulphur (kg/min)	Water (kg/min)	Sulphur (kg/min)	Water (kg/min)	Water (kg/min)	Lime (kg/min)	Solids (kg/min as FA & hyd. solids)	Water (kg/min incl. H_2O of hyd.)	Solids (kg/min as FA & hyd. solids)	Water (kg/min incl. H_2O of hyd.)
#3	30 968	24.8	3069	9.9	5037	2295.5	0	577	1363	577	327.5

NOTE: Coal Type #3 - Western Lignite, 0.5% S, AHV - 15 MJ/kg

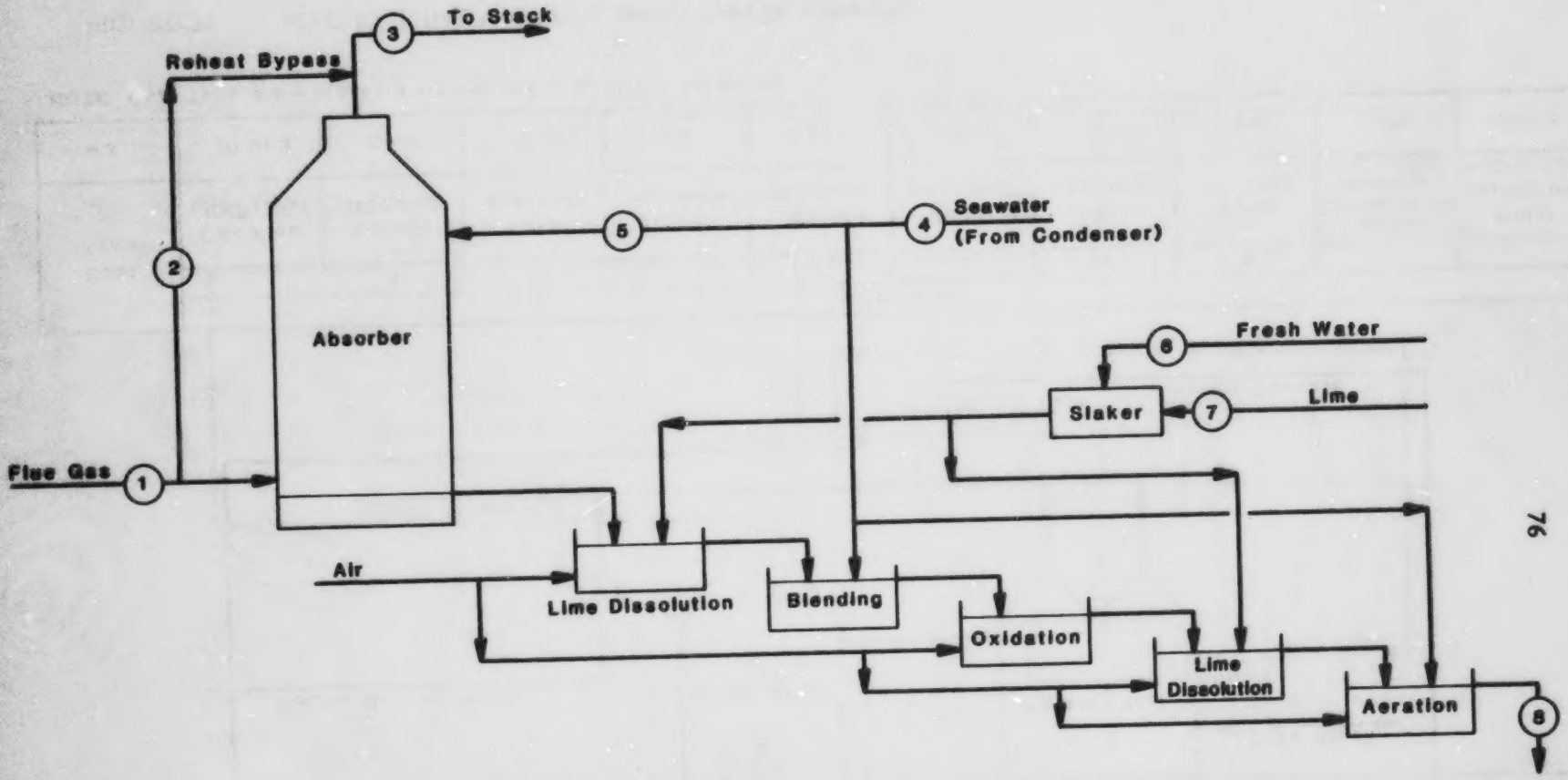
FIGURE 19 MATERIAL BALANCE FOR LIME-ALKALINE FLY ASH PROCESS



Coal Type	STREAM										
	1			2		3	4	5	6	
	Gas Flow (Nm^3/min)	Sulphur (kg/min)	Water (kg/min)	Sulphur (kg/min)	Water (kg/min)	Water (kg/min)	Lime (kg/min)	Water (kg/min)	Solids (kg/min as FA, inerts & hyd. solids)	Water (kg/min incl. H_2O of hyd.)
#3	30 966	24.8	3069	9.9	4852	1892	14	102	578	110.4

NOTE: Coal Type #3 - Western Lignite, 0.5% S, AHV - 15 MJ/kg

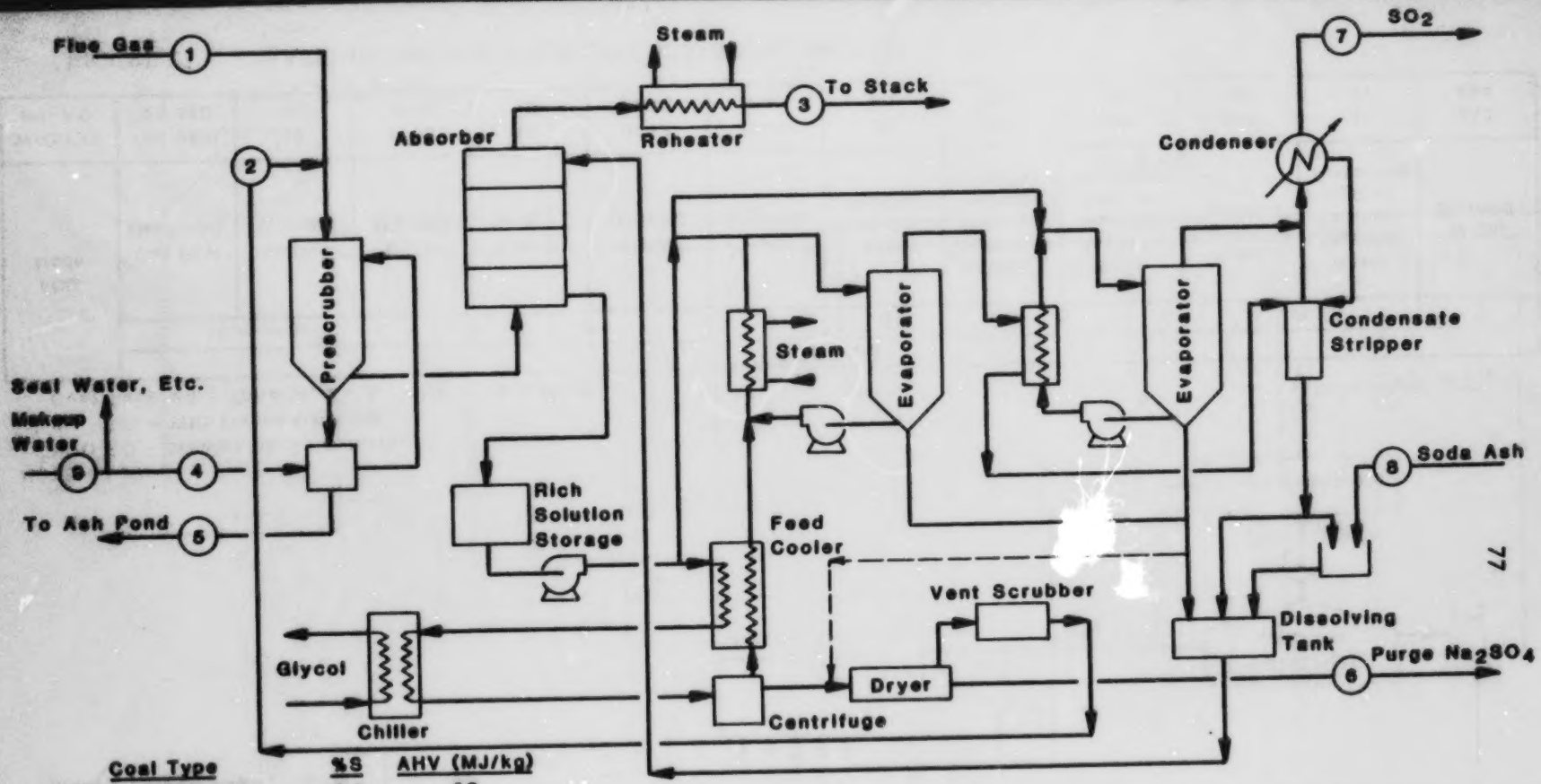
FIGURE 20 MATERIAL BALANCE FOR SPRAY DRYER PROCESS



NOTE: Coal Type #2 – Med. S Bituminous, 4.0% S AHV – 28 MJ/kg

Coal Type	STREAM 1			2			3		4	5	6	7	8
	Gas Flow (Nm^3/min)	Sulphur (kg/min)	Water (kg/min)	Gas Flow (Nm^3/min)	Sulphur (kg/min)	Water (kg/min)	Sulphur (kg/min)	Water (kg/min)	Water (kg/min)	Water (kg/min)	Water (kg/min)	Lime (kg/min)	Water (kg/min)
#2	22 767	99.3	1555	1821.4	7.9	124	9.9	715	837 500	167 500	245	132	838 586

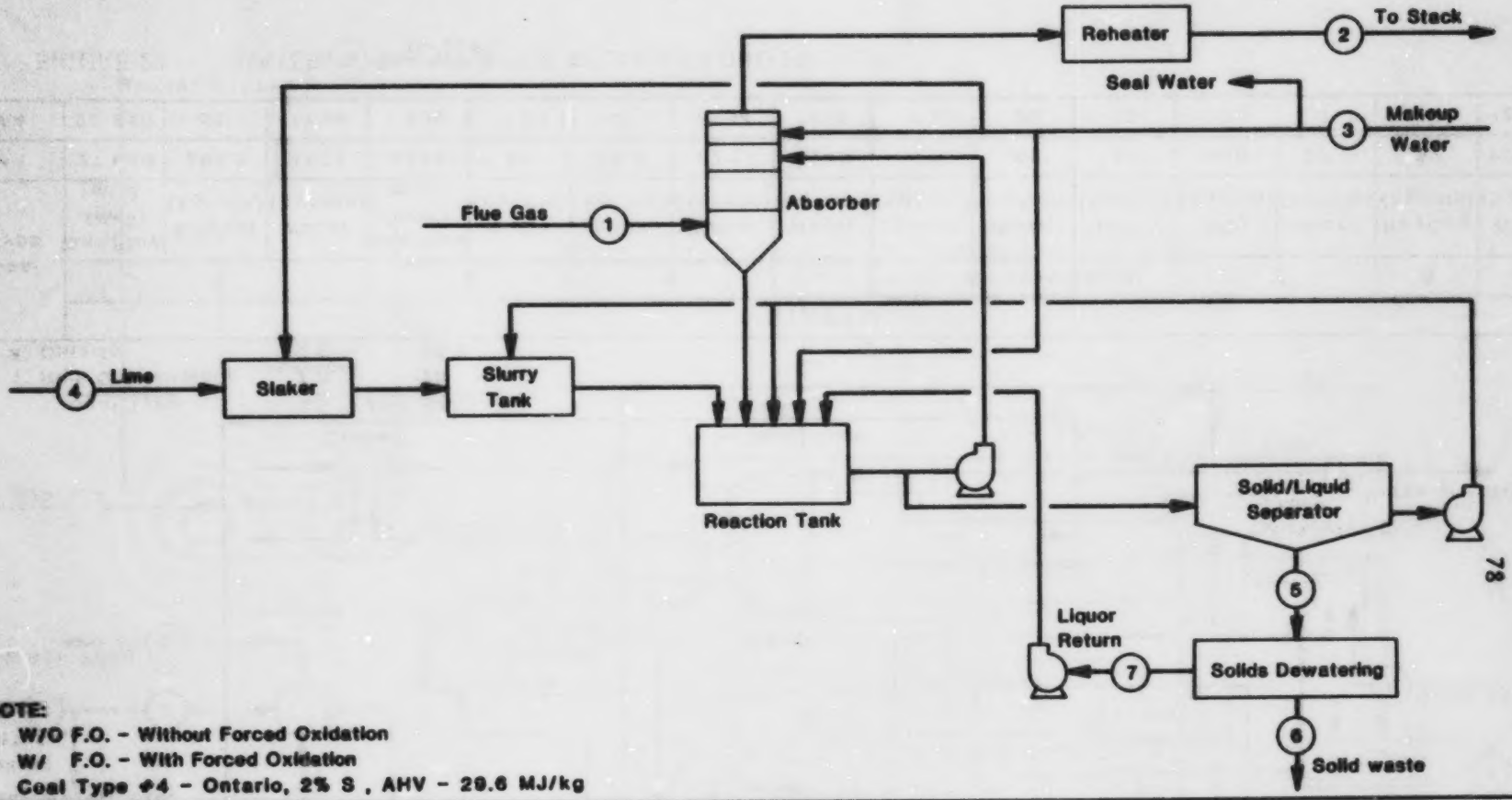
FIGURE 21 MATERIAL BALANCE FOR SEAWATER SCRUBBER PROCESS



Coal Type	%S	AHV (MJ/kg)
1. High S Bituminous	7.0	25
4. Ontario	2.0	29.6

Coal Type	STREAM														
	1			2		3		4	5		6	7		8	9
	Gas Flow (Nm^3/min)	Sulphur (kg/min)	Water (kg/min)	Gas Flow (Nm^3/min)	Water (kg/min)	Sulphur (kg/min)	Water (kg/min)	Water (kg/min)	Solids (kg/min)	Water (kg/min)	Solids (kg/min)	SO_2 (kg/min)	Water (kg/min)	Na_2Co_3 (kg/min)	Water (kg/min)
#1	27 630	208.8	1801	3098	54	20.9	3711	1948	4.9	92	81	338	37.5	62	2039.5
#4	22 250	52	1739	330	12	10	3294	1576	1.7	33	18	75	8	14	1606

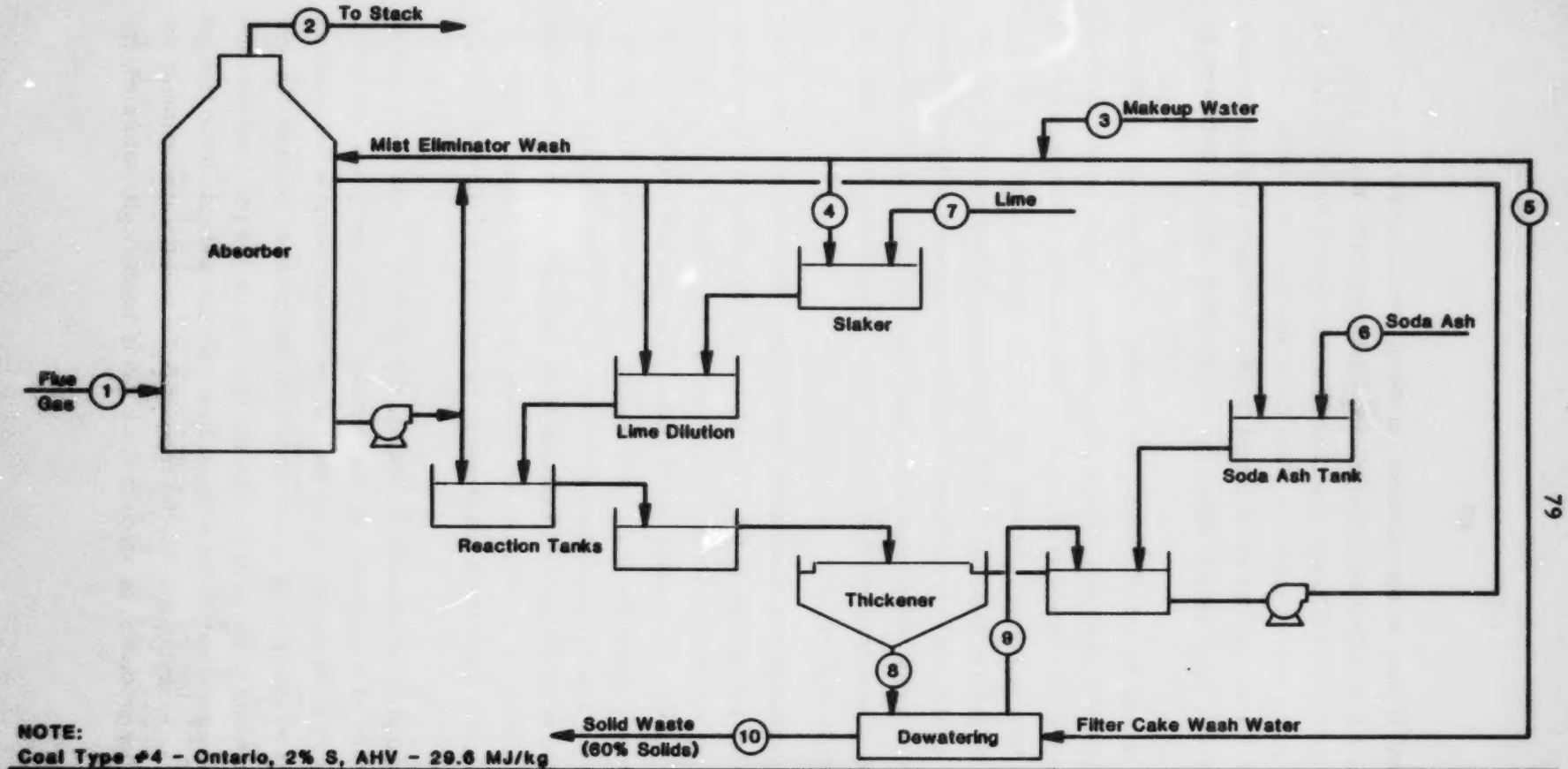
FIGURE 22 MATERIAL BALANCE FOR WELLMAN-LORD PROCESS



NOTE:
W/O F.O. - Without Forced Oxidation
W/ F.O. - With Forced Oxidation
Coal Type #4 - Ontario, 2% S , AHV - 29.6 MJ/kg

FGD Mode	STREAM											
	1			2		3	4	5		6		7
	Gas Flow (Nm^3/min)	Sulphur (kg/min)	Water (kg/min)	Sulphur (kg/min)	Water (kg/min)	Water (kg/min)	Lime (95% CaO) (kg/min)	Solids (kg/min as $CaCO_3$ & hydrated $CaSO_x$)	Water (kg/min including H_2O of hydration)	Solids (kg/min as hydrated $CaSO_x$)	Water (kg/min including H_2O of hydration)	Water (kg/min)
W/O F.O.	25 250	52	1739	10	3282	1684	80	187	454	187	141	313
W/ F.O.	25 250	52	1739	10	3282	1640	80	222	561	222	97	464

FIGURE 23 MATERIAL BALANCE FOR LIME SCRUBBING PROCESS



NOTE:
Coal Type #4 - Ontario, 2% S, AHV - 29.6 MJ/kg

Coal Type	STREAM														
	1			2		3	4	5	6	7	8		9	10	
	Gas Flow (Nm^3/min)	Sulphur (kg/min)	Water (kg/min)	Sulphur (kg/min)	Water (kg/min)	Water (kg/min)	Water (kg/min)	Water (kg/min)	Soda Ash (kg/min)	Lime 95% CaO (kg/min)	Solids (kg/min $CaCO_3$ & hydrated $CaSO_X$)	Water (kg/min)	Water (kg/min)	Solids (kg/min $CaCO_3$ & hydrated $CaSO_X$)	Water (kg/min)
#4	22 260	52	1739	10	3282	1685	220	250	7	80	187	454	313	187	141

FIGURE 24 MATERIAL BALANCE FOR DUAL ALKALI SCRUBBING PROCESS

Reagent lime is used in the seawater scrubbing process purely for scrubber liquor neutralization, not SO_2 absorption. The process vendor indicates that a stoichiometric ratio of 0.8 moles Ca/mole SO_2 removed is sufficient to restore the pH to 7.0, due to the natural alkalinity available in seawater. This lower stoichiometric ratio, coupled with a lower equivalent weight of CaO as compared to $CaCO_3$, results in a substantially lower reagent requirement, on a weight basis, for the seawater process compared to limestone scrubbing.

Soda ash makeup to the Wellman-Lord process is required to replace the portion of the sodium sulphite absorbent that is oxidized to unreactive sodium sulphate and is purged from the system. Attempts to reduce sodium losses by adding antioxidants to the scrubbing liquor have reportedly met with only limited success. The cost of using antioxidant is apparently greater than the cost of replacing the sodium purged as sodium sulphate. The use of chilled wall crystallizers to selectively remove sodium sulphate without encountering high sodium sulphite losses has been more successful (2, 25). For the purposes of this study, it was assumed that sodium losses would necessitate soda ash makeup at a stoichiometric ratio of 0.10.

6.2.3 Water Requirements and Wastewater Generation. Water is lost from FGD systems primarily through evaporation in the flue gas, through blowdown in the scrubber and through occlusion in the solid waste or entrainment in the product stream. Water must be added to the process to replace these losses. The primary points of water addition include reagent preparation (e.g., lime slaking and limestone milling) mist eliminator wash, pump seals, and in some cases waste sludge filter cake washing. Makeup water may also be added to various process tanks for level control, when makeup through the above points of addition is insufficient to replace the water losses. Table 27 summarizes the water losses and makeup requirements for each of the FGD processes and coal types considered.

With the exception of seawater scrubbing, the majority of the water loss in FGD systems is a result of evaporation in the scrubber. For the cases examined, evaporation represents 72 to 98 percent of the total water consumption. In the seawater scrubbing case, no net evaporation of water occurs in the scrubber because the low temperature of the scrubbing liquor cools the flue gas to below its dew point and moisture within the gas is condensed and adds to the effluent flow from the process. Since flue gas is not saturated in the spray drying process, this represents the lowest water consumer for a given coal. In all other processes, in which the flue gas is adiabatically saturated, the

TABLE 27 WATER REQUIREMENTS FOR FGD PROCESSES[a]

FGD Process	Coal Type[b]	Losses (L/min)					Makeup (L/min)		
		Evaporation in Scrubber	Occlusion in Solid Waste[c]	Entrainment in Product Stream	Blowdown	Total	Reagent Preparation	Other[d]	Total
Limestone									
Without Forced Oxidation	1	1856	697	-	-	2 553	1318	1 235	2 553
	2	1518	330	-	-	1 848	624	1 224	1 848
	3	1968	58	-	-	2 026	110	1 916	2 026
	4	1543	155	-	-	1 698	293	1 405	1 698
With Forced Oxidation	1	1856	464	-	-	2 320	1318	1 002	2 320
	2	1518	220	-	-	1 738	624	1 114	1 738
	3	1968	39	-	-	2 007	110	1 897	2 007
	4	1543	103	-	-	1 646	293	1 353	1 646
Spray Dryer	3	1782	110	-	-	1 892	26	1 866	1 892
Alkaline Fly Ash	3	1968	328	-	-	2 296	64	2 232	2 296
Seawater	2	(840)[e]	-	-	838 585	837 745	245	837 500	837 745
Wellman-Lord	1	1856	54[f]	38	92	2 040	-	2 040	2 040
	4	1543	12[f]	18	33	1 606	-	1 606	1 606
Lime									
Without Forced Oxidation	4	1543	141	-	-	1 684	220	1 464	1 684
With Forced Oxidation	4	1543	97	-	-	1 640	220	1 420	1 640
Dual Alkali	4	1543	141	-	-	1 684	220	1 464	1 684

a See Table 26 for assumptions used to calculate water losses and makeup requirements.

b

Coal Types		%S	AHV (MJ/kg)
1)	High S - Bituminous	7.0	25
2)	Med. S - Bituminous	4.0	28
3)	Western Lignite	0.5	15
4)	Ontario	2.0	29.6

c Includes water of hydration.
d Calculated by difference (i.e., total water losses - reagent preparation requirements). Includes mist eliminator wash, seal water, etc.
e Moisture condensed from flue gas.
f Water evaporated in purge solids dryer.

magnitude of the evaporative loss in the scrubber is the same for a given coal, assuming the same inlet temperature.

Occluded water in the solid waste represents the second largest loss from most processes. Water occluded in the solid waste of the lime, limestone, alkaline fly ash and dual alkali processes consists of both water of hydration and free water associated with the dewatered sludge. Water of hydration is chemically bound within the crystalline structure of the sludge, whereas free water is physically entrained interstitial water which can be released through dewatering and drying. To calculate the quantity of water occluded in the solid waste of the lime and limestone processes, it was assumed that 15 percent oxidation of sulphite to sulphate would occur if the system was not operated in a forced oxidation mode and 85 percent oxidation would be achieved with forced oxidation. Since most alkaline fly ash processes experience nearly complete oxidation, it was assumed that 100 percent of the sulphur in the solid waste from this process would be present as sulphate.

It was further assumed that all calcium sulphite would occur as the hemihydrate (i.e., $CaSO_3 \cdot 1/2H_2O$) and calcium sulphate would occur as gypsum (i.e., $CaSO_4 \cdot 2H_2O$). As gypsum sludges exhibit much better dewatering characteristics than sulphite sludges, it was assumed that the sludges from the lime and limestone forced oxidation processes would dewater to 80 percent hydrated solids, whereas only 60 percent solids would be achieved without forced oxidation (26). While the reaction products from the alkaline fly ash process largely comprise gypsum, the presence of large quantities of fly ash apparently detracts from the dewatering properties of the waste sludge. It was assumed that sludge from this process would only dewater to 65 percent solids.

Since the seawater scrubbing process is operated in a once-through manner, the scrubber blowdown stream comprises the total water loss from the system.

In the Wellman-Lord process, water occluded in the sodium sulphate purge is evaporated in a dryer. The water loss associated with drying the purge solids was calculated by assuming the wet cake fed to the dryer from the centrifuge would contain 60 percent solids and would be dried to 100 percent. Other water losses associated with the Wellman-Lord process include water entrained in the SO_2 product stream, which was assumed to be 10 percent, and the prescrubber blowdown. It was assumed that blowdown would be required to maintain a maximum suspended solids concentration of 5 percent by weight in the prescrubber loop. In practice, chloride concentration may be the limiting parameter, particularly with high chloride coals. It is normally necessary to keep the

chloride concentration in the prescrubber loop below 10 000 - 20 000 mg/L for corrosion protection (27).

To estimate water requirements for reagent preparation, it was assumed that lime would be slaked and limestone would be ground to a 35 percent solids consistency using makeup water. It was assumed that dissolution of soda ash and further reagent dilution would be accomplished using recycled scrubber liquor. Several facilities also use scrubber liquor for limestone milling, provided dissolved solids concentrations do not lead to equipment corrosion problems. This could conceivably reduce the fresh water makeup requirements for reagent preparation beyond the values indicated in Table 27.

Because of the considerable variation between facilities, depending upon design, no attempt was made to estimate the magnitude of other specific makeup water requirements. Seal water and mist eliminator wash can both be substantial consumers and contribute significantly to the overall water balance. Likewise, if filter cake washing is required for reagent recovery or to permit waste utilization, this too can demand significant makeup water. The "Other Makeup" column in Table 27 indicates the maximum quantity of fresh water that can be added to the process for these operations while still maintaining a closed water loop.

While the Wellman-Lord process typically operates with a blowdown stream from the prescrubber for control of suspended solids and chlorides, current trends in the design of most lime, limestone, alkaline fly ash and dual alkali processes are directed towards closed-loop operation with respect to the water balance. Nonetheless, a water imbalance can be created by any of several upset conditions, which may necessitate a blowdown. For instance, under low power plant load conditions, boiler flue gas flow is reduced, thereby reducing the amount of water lost to evaporation and occlusion in the solid waste. However, the amount of frest water makeup required by the process for pump seals and demister washing does not decrease proportionately. This may create a water imbalance and necessitate a purge.

Imbalances can also be caused by excessive fresh water intrusion into the system under normal operating conditions. Lack of maintenance on pump seals, use of large quantities of fresh water for FGD area washdown, and seasonally high inputs of rainfall or runoff can exceed the sytem's capacity to accept the necessary process inputs of fresh water without a blowdown.

Similarly, a change to a lower sulphur coal or increases in the extent of oxidation from sulphite to sulphate caused by changes in excess air in the flue gas may decrease sludge quantities produced or alter sludge dewatering characteristics sufficiently

to reduce the amount of water removed with the solids and necessitate a rapid blowdown of scrubbing liquor. Normal piping and seal leaks and equipment washdowns will also generate wastewaters.

The seawater scrubbing process creates a major wastewater stream, approximately equivalent to the station cooling water flow. The spray dryer process on the other hand produces no liquid effluent.

6.2.4 Solid Waste. Table 28 summarizes the quantities of FGD sludge and ash which would be generated by a hypothetical 500-MW station over a 20-year period, using the design criteria presented earlier. While forced oxidation of lime and limestone sludges produces approximately 20 percent more dry solids (by weight) than are contained in predominantly sulphite sludges, the total volume of wet FGD sludge ultimately requiring disposal is actually about 25-30 percent less, assuming mechanical dewatering and landfill of both materials. If ponding is used for the sulphite sludge, as is common practice, the required disposal volume would be considerably greater.

The dual alkali process would produce approximately the same quantity of sludge as a lime FGD system, without forced oxidation, assuming it is operated at the same stoichiometric ratio of calcium to sulphur dioxide.

The seawater scrubbing process produces no solid waste, as such, because sufficient water is added to keep all sulphur reaction products in solution. However, residual fly ash scrubbed in the absorber and inerts added with the lime for neutralization are discharged with the scrubber effluent as suspended solids.

The total volume of solid waste produced by spray drying in the western lignite case would be about 30 percent less than if the alkaline fly ash process were used, and approximately 15-20 percent less than if limestone scrubbing were used. This is due primarily to the lower water content of the spray dryer solids, as disposed.

The Wellman-Lord process would produce only a fraction of the quantity of waste solids generated by any of the throwaway processes, which represents a significant environmental advantage.

6.3 Process Design Alternatives

Several process design options exist for each of the FGD processes considered in this study. Many of these design options affect the amount of fresh makeup water required by the process, the need for blowdown of liquid waste from the FGD system, and the quantity and characteristics of solid waste produced. The process design alternatives

TABLE 28 SOLID WASTE PRODUCTION RATES AND DISPOSAL VOLUMES[a]

FGD Process	Coal Type[b]	Solid Waste Production (kg/min)				20 Year Landfill Volume (10^6 m^3)			
		FGD Solids[c] (dry)	FGD Sludge[d] (as disposed)	Bottom Ash (dry)	Fly Ash[e] (dry)	FGD Sludge[f]	Bottom Ash[g]	Fly Ash[g]	Total
Limestone									
Without Forced Oxidation	1	931	1551	162	486	6.89	0 77	2.31	9.97
	2	441	735	81	243	3.07	0.38	1.14	4.59
	3	78	130	175.3	526	0.58	0.83	2.49	3.90
	4	207	345	58.3	175	1.53	0.28	0.84	2.65
With Forced Oxidation	1	1107	1384	162	486	5.04	0.77	2.31	8.12
	2	524	656	81	243	2.39	0.38	1.14	3.91
	3	93	116	175.3	526	0.42	0.83	2.49	3.74
	4	247	309	58.3	175	1.12	0.28	0.84	2.24
Alkaline Fly Ash	3	577[h]	888	175.3	-	3.77	0.83	-	4.60
Spray Dryer	3	578[h]	680	175.3	-	2.35	0.83	-	3.18
Seawater	2	-	-	81	243	-	0.38	1.14	1.52
Wellman-Lord	1	81	81	162	486	0.23	0.77	2.31	3.31
	4	18	18	58.3	175	0.051	0.28	0.84	1.17
Lime									
Without Forced Oxidation	4	187	312	58.3	175	1.38	0.28	0.84	2.50
With Forced Oxidation	4	222	278	58.3	175	1.01	0.28	0.84	2.13
Dual Alkali	4	187	312	58.3	175	1.38	0.28	0.84	2.50

[a] For hypothetical 500-MW station operating at 37 percent energy conversion efficiency, 65 percent load factor.

[b]

Coal Type		%S	AHV (MJ/kg)
1)	High S Bituminous	7.0	25
2)	Med. S Bituminous	4.0	28
3)	Western Lignite	0.5	15
4)	Ontario	2.0	29.6

[c] See Table 26 for assumptions used to calculate FGD solids production rates. FGD solids include water of hydration.

[d] Limestone and lime sludge without forced oxidation is assumed to contain 60 percent solids. Limestone and lime sludge with forced oxidation is assumed to contain 80 percent solids.

[e] Bottom ash/fly ash split = 25 percent/75 percent.

[f] Specific gravity of 2.40 g/cm^3 was assumed for all dry FGD solids.

[g] Dry disposal of bottom ash and fly ash assumed. Dry bulk density was assumed to be 1.44 g/cm^3.

[h] FGD solids include fly ash.

that have demonstrated a significant effect upon the FGD system material balance, waste characteristics or performance are described below.

6.3.1 Cascade Strategies for Water Reuse. Most FGD systems are designed for a high degree of water reuse within the process. Scrubber liquor extracted from sludges during thickening and dewatering is normally returned to the scrubber loop to minimize fresh water makeup requirements and wastewater blowdown. Some utilities have taken this philosophy one step further and have attempted to use other wastewater streams within the overall station water system as the source of makeup to the FGD system. Waste process streams that have been used to provide makeup water to FGD systems include cooling tower blowdown, coal pile runoff, water treatment plant wastes (e.g., filter backwash and demineralizer rinse water), ash handling system blowdown, and miscellaneous sump and drain wastes. Depending upon the desired reuse application, and the wastewater quality, pretreatment of these streams may or may not be required prior to reuse. To evaluate the potential for wastewater reuse, it is useful to examine the makeup water quality requirements of the principal water consumers.

The quality of the water used for slaking lime is probably more critical than for any other scrubber use and should be close to potable quality. Slaking is an exothermic hydration reaction whereby CaO is converted to $Ca(OH)_2$. Waste or recycle process waters containing sulphites and sulphates retard slaking and reduce the quality of the lime slurry. Poor quality slaking water causes a larger average slaked particle size. The smaller surface area reduces reagent reactivity in the scrubber. In fact, some of the lime does not hydrate and is wasted. It appears that the lime precipitates the SO_3 and SO_4 ions as calcium sulphite/sulphate, which coats the unreacted CaO particles and prevents the complete penetration of water. Slaking water should be free of high concentrations of such ions as carbonates, bicarbonates, sulphates, or phosphates that will precipitate in the presence of calcium and cause a scaling problem. Other metal ions that will precipitate in the presence of hydroxide ions are also objectionable. High concentrations of chlorides in slaking water do not appear to be detrimental to the slaking process; however, they may contribute to equipment corrosion (28).

Opinions differ regarding the use of recycled scrubber water for slurry dilution. If the lime slaker is operated properly and if complete slaking of quicklime has been achieved, dilution with recycled water should be satisfactory, and only minimal use of fresh water would be recommended. In fact, the addition of magnesium or sodium ions to the scrubber liquor is considered to have a beneficial effect upon SO_2 removal and to aid in the reduction of scaling (28).

The preparation of limestone reagent does not involve slaking or hydration. Rather, it is purely a mechanical wet grinding operation normally carried out in ball mills. Recycled scrubber liquor is commonly used both in the grinding operation and for slurry dilution, provided dissolved solids concentrations do not create equipment corrosion or scaling problems. Likewise, recycled scrubber liquor is suitable and commonly used for soda ash preparation.

Plugging and scaling are frequently encountered mist eliminator problems in calcium-based scrubbers. A soft, mudlike deposit and/or scale can accumulate on the mist eliminator in the course of time, unless it is sprayed with sufficient wash water of suitable quality. If solids build up to the point where the collector is completely blocked and "blow holes" develop, the result is increased pressure drop across the mist eliminator and reduction in efficiency (28).

Solid deposits of calcium sulphite and unreacted lime can occur in lime systems when the solid carryover from the scrubber is trapped in the mist eliminator. More serious, however, is formation of sulphate scale due to oxidation of the sulphite solution collected on the mist eliminators. Scaling can also occur as the result of absorption of residual flue gas SO_2 by the unreacted lime on wetted surfaces. Surface irregularities formed by the crystalline scale increase the potential for mud accumulation and decrease the effectiveness of washing operations.

Fresh water is normally preferred for mist eliminator washing. However, to maintain a closed-loop water balance, 100 percent fresh water washing is often not possible. The normal procedure in calcium-based scrubbing systems is to introduce all makeup fresh water (other than that required for pump seals and lime slaking) through the mist eliminator wash system. Additional wash water, which is sometimes required, is usually obtained by recycling clear water from the thickener or from the sludge pond overflow. The disadvantage of using this liquor is that it is frequently saturated with calcium sulphate. If calcium carried up into the mist eliminator reacts with residual sulphur dioxide in the gas to form more calcium sulphate, then precipitation and scaling can occur (28).

Use of recycled scrubber liquor for mist eliminator wash has met with varied results. Some utilities have reported successful operation using 100% recycled water, whereas others have encountered scaling problems. Several use a blend of fresh water and recycle liquor to de-saturate the wash water with respect to calcium sulphate; some vendors recommend using process water sequentially with fresh water, rather than

blending. An alternative, though more costly, approach would be the treatment of recycle liquor with soda ash to remove calcium and permit a higher rate of reuse.

The volumes of water required for mist eliminator washing are substantial and are frequently the greatest single obstacle to closed-loop operation, particularly when a large proportion of fresh water is used. Efforts have been directed not only towards increasing the use of recycle liquor, but towards decreasing the volumetric wash water requirements. Short duration, high pressure washes for portions of the mist eliminator rather than continuous washing of the entire surface have been used. Some mist eliminator designs are also easier to keep clean than others.

Yet another approach is the separation of the mist eliminator washing loop from the scrubbing loop. By installing a tray beneath the mist eliminator, the wash water can be captured and recycled without adding significantly to the water balance in the absorber. Fresh makeup water is required to replenish wash water which is sent to the absorber to maintain dissolved solids below the saturation concentration of scaling salts in the wash water loop. However, the size of this stream is significant'y smaller than the volume of fresh water typically added to the system with unsegregated washing and scrubbing loops. The obvious disadvantage of separate loops is the additional cost and complexity of the system.

Pump seal water is normally required to contain no particulate matter larger than 40 µm and to have a non-scaling index (28). Fresh water is most frequently recommended because the heat associated with pump operation can cause scale deposition and consequent pump failure if scrubbing liquor saturated in calcium sulphate and other salts is used. Some facilities have reused process water which has been filtered and diluted with fresh water, but 'imited experience with this approach has been reported. As seal water becomes part of the total water inventory in the FGD system, the use of excessive quant ties of fresh water for this purpose reduces the amount that can be used for other operations (e.g., mist eliminator washing) while maintaining a closed water loop.

Waste sludge from the dual alkali process is normally washed during dewatering to recover sodium. A water supply with relatively low dissolved solids content is recommended for this applicat on, and fresh water is most commonly used. Forced oxidation systems designed to produce by-product gypsum may also require filter cake washing with fresh water to remove chlorides from the waste product.

The quality of water is not normally as critical for other FGD requirements, such as makeup for level control to process tanks. These requirements provide the greatest potential for reuse of waste process streams and recycled scrubber liquor.

However, in contemplating any wastewater reuse scheme, due consideration must be given to the possible effects upon the FGD system. For instance, streams which significantly alter the scrubbing liquor pH may have an undesirable effect upon SO_2 removal efficiency or absorbent utilization. Likewise, streams containing high dissolved solids concentrations, particularly chlorides and fluorides, may create equipment corrosion problems or may contaminate an otherwise saleable by-product (e.g., gypsum). Flow rates and flow frequencies also must be carefully considered in the overall water balance. For example, coal pile runoff is usually a seasonal source of water, and demineralizer wastes are only intermittently available. Thus adequate storage and flow balancing provisions are required to successfully implement cascade strategies for water reuse.

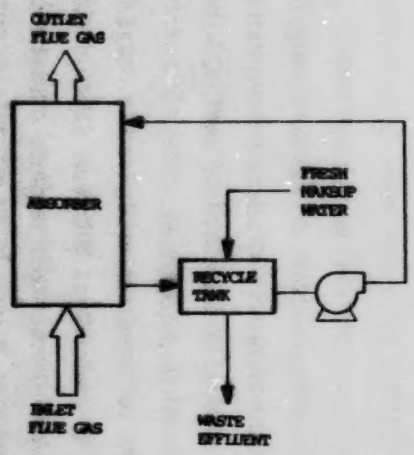
6.3.2 **Double-Loop Operation.** Maintaining the water balance in a closed-loop scrubbing system necessitates reusing all return water from sludge thickening and dewatering operations as well as accepting some minimum addition of relatively high quality makeup water to the process for the more critical requirements such as pump seal water, demister washing and reagent preparation. Due to the buildup of dissolved salts that results from closed-loop operations, calcium sulphate scaling and chloride-induced corrosion are two problems that have commonly occurred in scrubbers operating with zero process liquid waste discharge.

Single-loop and double-loop scrubbing systems are illustrated schematically in Figure 25. The advantages claimed for double-loop design of limestone scrubbers include the following (28):

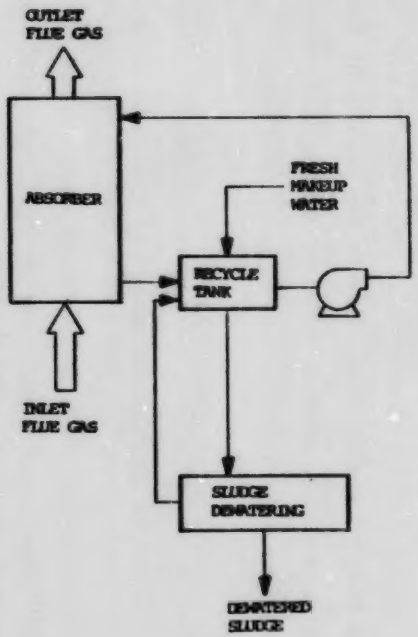
a) better water balance control,
b) better control of scaling and corrosion in the absorber,
c) high reagent utilization with high SO_2 removal efficiency, and
d) high rates of sulphite oxidation to sulphate.

This process modification uses two distinct scrubbing loops; one in the absorber, the other in the quencher or prescrubber section of the scrubbing tower. The absorber loop operates at a higher pH and lower dissolved solids concentration than the quencher loop. Open-loop operation is maintained in the absorber section with blowdown of partially spent liquor being sent to the closed-loop quencher.

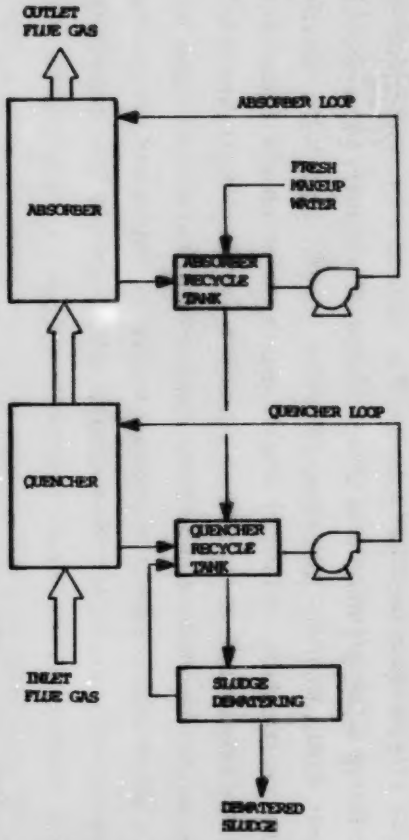
Fresh water enters the absorber as demister wash water, reagent slurrying water and as pump seal water. The pH is maintained between 6.0 and 6.5, which is optimum for high levels of SO_2 absorption but discourages sulphite oxidation. Thus the



(A) SINGLE LOOP SCRUBBING IN OPEN-LOOP CONFIGURATION



(B) SINGLE LOOP SCRUBBING IN CLOSED-LOOP CONFIGURATION



(C) DOUBLE LOOP SCRUBBING

FIGURE 25 SINGLE-LOOP AND DOUBLE-LOOP SCRUBBING SYSTEMS

absorber loop contains low levels of chlorides and is not saturated with calcium sulphate, resulting in a relatively non-scaling, non-corrosive environment.

The quencher loop accepts primarily blowdown from the absorber and return water from sludge dewatering operations. Fresh water can also be added if additional makeup is required. The closed-loop operation of the quencher results in a high concentration of dissolved solids. It is operated at a lower pH (approximately 5) which promotes high limestone utilization and oxidation of sulphite to sulphate. Although SO_2 absorption is relatively low at the depressed pH, undissolved and unreacted limestone blowdown from the absorber loop is dissolved and utilized. This section is operated at supersaturated levels of calcium sulphate causing the precipitation of gypsum crystals and better dewatering and handling characteristics than calcium sulphite sludge. Air can be added to the quencher loop, if desired, to ensure full oxidation.

Double-loop operation offers high SO_2 removals and high reagent utilization, two apparently conflicting objectives in single-loop systems because of the different pH environments required for each. Thus the quantity of limestone required and solid waste produced by a double-loop system would be expected to be slightly less than for a single-loop system.

Since the high chloride, high sulphate environment is restricted to the closed-loop quencher section, which is normally a relatively simple venturi tower, the problems of scaling and corrosion of absorber section packing, spray nozzles, trays and demisters are largely avoided. However, because double-loop operation requires more slurry recirculation pumps than a single-loop configuration, seal water input to the system may increase.

6.3.3 Forced Oxidation. In early calcium-based scrubbers attempts were made to minimize the oxidation of sulphite to sulphate (gypsum), as it was believed this was essential to control scale formation. More recently, however, several lime and limestone FGD systems have been redesigned to encourage sulphite oxidation without the associated scaling problems. The advantages of increasing the sulphate to sulphite ratio are numerous:

a) Calcium sulphate has better dewatering properties.
b) Smaller sludge disposal volumes are required.
c) Calcium sulphate has better handling and structural properties, which facilitate landfill operations and disposal site reclamation.
d) Disposal of calcium sulphate is less of an environmental problem. Leachates from sulphite sludges can impose a high chemical oxygen demand.

e) There is a potential reuse market for the sulphate by-product in the manufacture of gypsum wallboard.

Sludges from calcium-based FGD systems operated without oxidation typically have undesirable properties when disposed without stabilization or fixation. They generally have poor adhesive strength and are difficult to dewater. The crystal structure is largely responsible for many of the undesirable characteristics of sulphite scrubber sludges.

Calcium sulphite crystals form extremely thin, fragile platelets, which usually occur in clusters or rosettes. The rosettes and platelets form an open structure, with water filling the voids. This occluded water is difficult to remove by gravity thickening or filtration. The rosettes are not easily compacted by normal settling, and filtering causes the fragile crystals to break, tending to clog the filter cloth. As a result, sulphite-predominant sludges will normally dewater to only about 30 to 35% solids by settling and can be filtered to about 50 to 60% solids (26). The crystal structure of sulphite sludges coupled with their water retention results in thixotropic behaviour, i.e., an apparently solid mass of sludge will reliquify when vibrated or disturbed. This characteristic virtually dictates that unstabilized sulphite sludges be ponded rather than landfilled because the material behaves more like a fluid than a solid.

The crystalline form of calcium sulphate is blocky and elongated, which results in better settling and dewatering characteristics than sulphite sludges. Sulphate sludges will settle to about 60 - 70 percent solids, and will vacuum filter to as much as 80 - 90 percent solids. They are easily handled by front-end loading equipment, can be hauled by truck or conveyor, and can be disposed of by stacking in structurally stable stock piles, some as high as 45 m.

As discussed in Section 6.2, whether or not a calcium-based FGD system is operated in a forced oxidation mode affects the quantity as well as the characteristics of solid waste produced. While more dry solids are actually produced (on a weight basis) with sulphate sludges than with sulphite sludges per mole of sulphur removed, the total amount of wet sludge ultimately to be disposed is normally less, due to the improved dewatering characteristics of the former.

As discussed in Chapter 3, several forced oxidation design options exist. These vary from simple single-loop systems in which air is sparged into the absorber reaction tank to double-loop systems such as those described in Section 6.3.2, to relatively complex catalyzed and buffered processes such as the Davy S-H process described in Section 3.8.3. Regardless of the process used, the conditions under which oxidation is enhanced are

similar. The optimum pH range is usually between 4.5 and 6.0, and air is typically required at a stoichiometric ratio of between 1.3 and 1.8 (26). The presence of some trace metals also appears to catalyze the oxidation reaction. Ferric chloride is added as a catalyst in both the Davy S-H and Chiyoda 101 processes. The presence of several trace metals in the scrubbing liquor in alkaline fly ash processes has also been cited as one reason for the typically high oxidation rates experienced with these systems.

One forced oxidation process of particular interest, because of its simplicity of design, is the Chiyoda Thoroughbred 121 (CT-121) process. As described in Section 3.2, SO_2 absorption, oxidation, and precipitation all occur in a patented gas/liquid contacting device referred to as a jet bubbling reactor. Some of the advantages claimed include the elimination of slurry recirculation pumps and spray nozzles, lower energy consumption, and low slurry entrainment in the gas stream, which reduces mist eliminator washing requirements. The process was successfully pilot tested at Gulf Power Company's Scholz Generating Station in 1978 and 1979.

While forced oxidation greatly improves the handling and disposal properties of solid waste generated by calcium-based scrubbers, it makes closed-loop operation, with respect to the water balance, significantly more difficult. The improved dewatering characteristics of gypsum, compared to sulphite sludges, means that more scrubbing liquor is returned to the process, reducing the effective blowdown rate from the system. This has two effects. First it reduces the amount of fresh makeup that can be added to the system for mist eliminator washing, seal water, reagent preparation, etc., while maintaining a closed water loop. Second, it increases the concentration of dissolved solids in the scrubbing liquor, reportedly as much as four to five times, creating a very aggressive corrosion environment. The opinion has been expressed that, even if forced oxidation systems could be volumetrically balanced to operate closed-loop, open-loop operation may be essential to limit corrosion effects to manageable levels (29).

The decision to include a forced oxidation step in a calcium-based FGD process is affected by several variables other than those discussed here, including cost, environmental considerations, and impact on process reliability, complexity, and operability.

6.3.4 Ponding vs. Dry Disposal of FGD Sludge. When ponding is used for FGD sludge disposal, climatic conditions can significantly affect the process water balance. For illustrative purposes, Table 29 presents the estimated average annual input of precipitation to the FGD system for some of the cases examined in this study. These estimates are

TABLE 29 AVERAGE ANNUAL PRECIPITATION INPUT TO SCRUBBER SLUDGE PONDS

Coal Type[a]	FGD Process	Location	Storage Volume[b] (10^6 m^3)	Pond Area[c] (10^3 m^2)	Net Precip.[d] (mm)	Annual Water Input Due to Precip. (L/min)	Annual Water Input Due to Precip. % Of Makeup
1	Limestone without forced oxidation	New Brunswick	5.21	347	600	396	15.5
2	Limestone without forced oxidation	Nova Scotia	2.47	165	1 000	314	17.0
3	Limestone without forced oxidation	Saskatchewan	0.44	29.3	20	1.1	0.05
4	Limestone without forced oxidation	Ontario	1.16	77.3	200	29.4	1.7
1	Alkaline Fly Ash	Saskatchewan	3.22[e]	215[e]	20	8.2	0.36

a

Coal Type		% S	AHV (MJ/kg)
1)	High S Bituminous	7.0	25
2)	Med. S Bituminous	4.0	28
3)	Western Lignite	0.5	15
4)	Ontario	2.0	29.6

b Storage volume based provision of sufficient capacity for 10 years sludge production at 45 percent solids content.

c Pond area based on 15 m pond depth.

d Net precipitation is the estimated annual water input due to precipitation minus evaporative losses and was estimated from derived annual runoff data presented in Tables 5 and 6.

e Pond size based on provision of storage capacity for all fly ash and scrubber sludge.

based on the assumption that ponds would be developed with sufficient storage capacity for 10 years sludge production at 45 percent solids content.

It is apparent that in eastern Canada precipitation input to sludge ponds could have a significant impact on the FGD system water balance. The larger ponds needed to accommodate the sludge production from high sulphur coals, coupled with the higher net precipitation, results in an estimated annual average precipitation input to the system roughly equivalent to 15 to 17 percent of the total makeup water requirements.

Because precipitation and evaporation are not evenly distributed over the year, it may be necessary to deal with a significantly higher percentage input of process water from rainfall or runoff on a seasonal basis. Since some minimum amount of fresh water is continuously required by the FGD system for pump seals, lime slaking, etc., the process may be incapable of absorbing this influx of excess water over a short duration, necessitating a blowdown.

Dry disposal of FGD sludge by landfill can greatly reduce the climate-created water imbalances in such cases. Dry sludge requires a significantly smaller disposal site and careful landfill design can minimize infiltration of precipitation into the site, greatly reducing or eliminating return flows that must be handled by the FGD system.

6.3.5 Process Surge Capacity. The amount of surge capacity provided within the FGD process can have a profound effect upon the system's ability to respond to transient water imbalances without blowdown of a wastewater stream. This is important both from an environmental perspective and from an economic one. Any release of scrubbing liquor represents a loss of reagent, which must ultimately be replaced. This is particularly critical in the case of sodium-based FGD systems (i.e., dual alkali and Wellman-Lord). These systems operate at a high concentration of dissolved solids with a high sulphite level, creating a high chemical oxygen demand. Discharge of this material to a watercourse would not only have adverse environmental implications, but the cost of soda ash replacement would be substantial.

Surge capacity can be provided by process tanks, sumps, or ponds. The recommended amount and type of surge capacity are very system dependent. In general sodium-based scrubbing systems should have greater surge capacity than calcium-based ones for the reasons described above. An emergency dump pond is advisable for these systems to enable scrubbing liquor to be recovered in the event of a major failure or the need to drain a process tank.

Sludge disposal sites such as ponds, landfills, and gypsum stacks should also have sufficient storage capacity to retain precipitation resulting from a reasonably conservative design storm.

6.3.6 Buffer Additives. Considerable research and development work in recent years has been directed towards improving the performance of limestone scrubbing systems. The addition of organic acid buffering agents to laboratory, pilot, and commercial-scale slurry scrubbers has demonstrated several significant advantages, including:

a) higher SO_2 removals,
b) increased limestone utilization,
c) greater system flexibility,
d) lower projected capital and operating costs, and
e) decreased waste sludge volume.

The absorption of SO_2 in limestone slurry scrubbing liquor is represented by the following equilibrium reaction:

$$SO_2 + H_2O \rightarrow H^+ + HSO_3^-$$

The drop in pH caused by the release of H^+ ions during absorption is buffered by the dissolution of limestone reagent as follows:

$$CaCO_3 + 2H^+ \rightarrow Ca^{++} + CO_2 + H_2O$$

The rate of SO_2 absorption decreases with decreasing pH, and thus is limited by the rate of $CaCO_3$ dissolution. Buffer additives (A^-) enhance SO_2 gas/liquid mass transfer by reacting reversibly with H^+ (an intermediate and fast reaction step), reducing the pH drop that normally occurs across the gas/liquid interface during SO_2 absorption, as follows:

$$A^- + H^+ \rightarrow HA$$

The resultant higher concentration of dissolved SO_2 in the liquid phase accelerates the rate of reaction to form $CaSO_3$ and $CaSO_4$. This has the overall effect of increasing SO_2 removal efficiency at a given liquid-to-gas ratio (L/G) and reaction tank detention time, or enabling equivalent SO_2 removals to be achieved with lower L/Gs and smaller reaction tanks.

Although several organic acids offer this buffering capacity, most work to date has been concerned with the use of adipic acid and dibasic acid (DBA). Adipic acid is

a raw material used in the manufacturing of nylon, whereas DBA is a less expensive waste by-product of adipic acid production. Figures 26 and 27 illustrate the dramatic effect of buffer additives on SO_2 removal and limestone utilization at a constant L/G. They also show that a number of sources of DBA give nearly identical results.

Preliminary economic evaluations indicate that addition of organic acid buffering agents is cost-effective. Although the additive cost is high (approximately $1.35/kg for adipic acid and $0.66/kg for DBA in 1981 U.S. $), this is offset by reduced limestone consumption, reduced grinding cost, and lower waste sludge disposal volumes (30, 31). The cost of using additive is primarily a function of the losses which must be made up, creating an economic incentive in buffer-enhanced limestone systems to maintain a tightly closed water loop and to dewater waste sludge to the maximum extent possible.

Investigations have shown that forced oxidation is not affected by the addition of adipic acid or DBA, and that no significant toxicity or other environmental problems appear to be introduced in the waste solids (30, 32). However, due to odours associated with degradation products of adipic acid and the set-retarding characteristics of carboxylic acid, the utilization potential of by-product gypsum may be adversely affected by the use of organic acid additives (33).

6.3.7 Regenerative Flue Gas Reheat. Regenerative flue gas reheating involves the transfer of sensible heat from the scrubber inlet gas to the saturated gas leaving the mist eliminators. Although the concept is in the early stages of commercial demonstration, preliminary ecomomic analyses indicate that it may offer substantial life-cycle savings in terms of reduced energy costs, compared to indirect reheating.

The reduction in raw gas inlet temperature which results from regenerative reheating can significantly reduce the evaporative water loss in the scrubber. This is considered a beneficial effect for FGD systems that derive their makeup from plant wastewater or other high TDS sources because of the corresponding reduction in cycles of concentration for makeup water dissolved solids. Reductions in FGD system cycles of concentration of from 25 to 60 percent have been reported for two facilities using regenerative reheat (34). Reducing the cycles of concentration reduces the dissolved solids concentration in the scrubbing liquor, which alleviates some corrosion problems and may lessen the environmental impact of a liquid discharge from the system.

Use of regenerative reheat in many facilities that have successfully operated closed-loop would necessitate implementation of a liquid purge, which otherwise may not

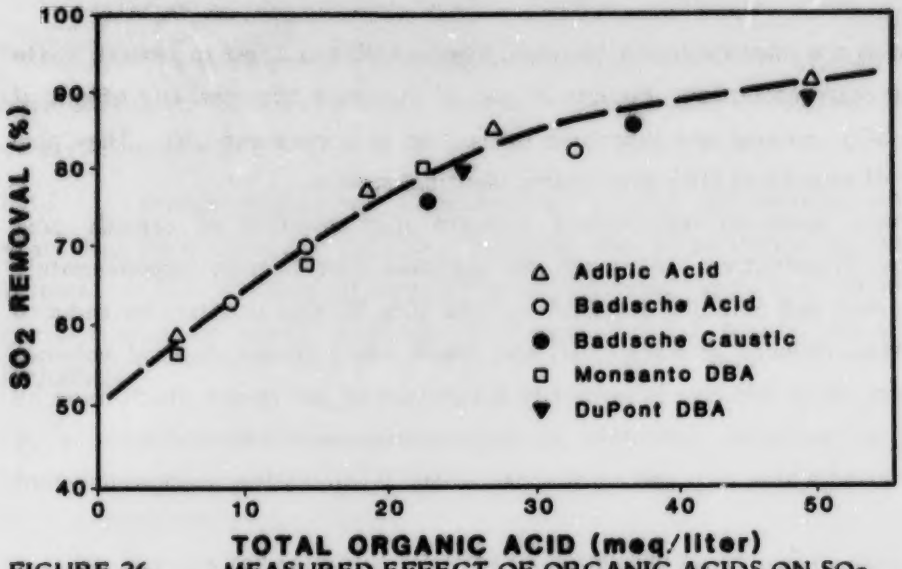


FIGURE 26 MEASURED EFFECT OF ORGANIC ACIDS ON SO_2 REMOVAL AT pH 5.0, L/G - 9 L/m^3 (30)

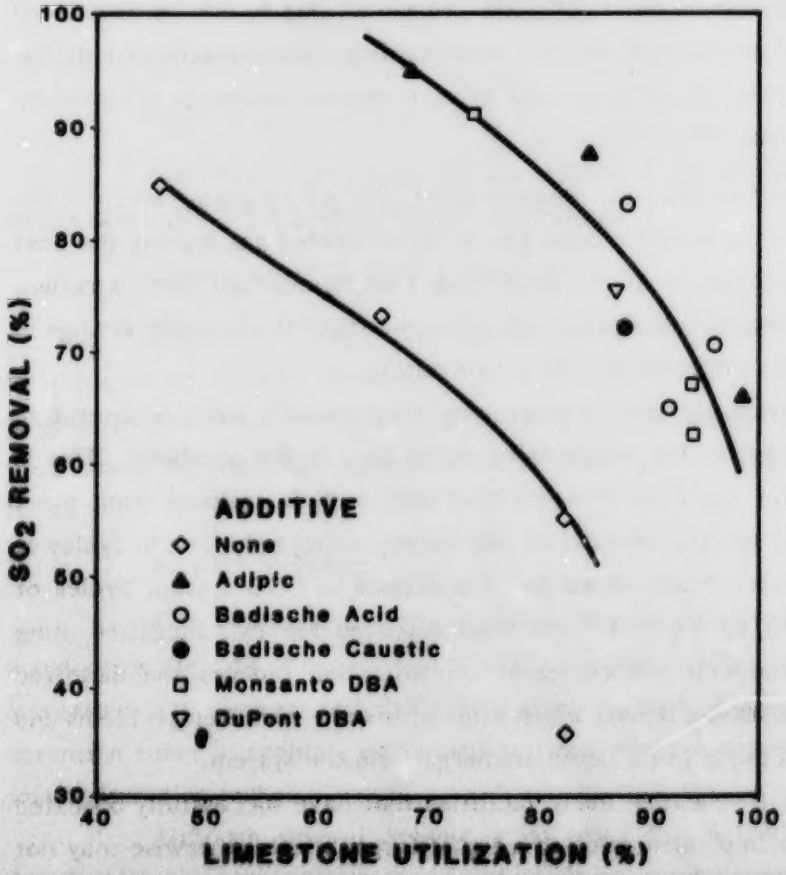


FIGURE 27 EFFECT OF 13.7 meq/L OF ORGANIC ACIDS ON LIMESTONE UTILIZATION AND SO_2 REMOVAL, L/G - 9 L/m^3 (30)

have been necessary. Most systems add some minimum quantity of fresh water for mist eliminator washing, pump seals, reagent preparation, etc. This water addition is offset primarily by evaporative losses. If, however, evaporation is significantly reduced, then makeup water will also have to be reduced to maintain the water balance and avoid a liquid purge. This can only be accomplished by substituting recycled process water for some of the makeup requirements.

6.3.8 Spray Dryer Solids Recycle. Initial pilot plant tests with spray dryer FGD systems were run on a single-pass basis and experienced relatively low reagent utilization, particularly at high SO_2 removal efficiencies. However, more recent test results with the recycle of solids captured in particulate removal equipment downstream of spray dryers have indicated that this design option can significantly reduce reagent requirements and presumably the quantity of solid waste produced by the process.

In one experimental program, alkaline fly ash from a North Dakota lignite, containing 23.8 percent CaO, was slurried and used as the sole source of spray dryer reagent. SO_2 removals of 75 percent and 71 percent were experienced when scrubbing flue gas from the combustion of a 3.1 percent sulphur West Virginia coal and a 1.0 percent sulphur Kentucky coal, respectively, at a Ca/S ratio of 1.2 and an 11°C approach to saturation temperature. The results of this test suggested that, when burning low sulphur lignite (10 percent ash, 0.8 percent sulphur), SO_2 removals of 70 percent could be achieved by slurrying approximately 70 percent of the fly ash and feeding it to a spray dryer, without any additional reagent (34).

Other pilot and commercial scale tests with solids recirculation have also indicated that recirculation improves reagent utilization by three mechanisms (14):

a) Unreacted sorbent is reintroduced to the spray dryer.
b) Available alkali in the fly ash is utilized.
c) Inert particles become coated with fresh reagent and provide a larger surface area for the absorption reactions to occur.

While only limited test data are available to date, preliminary results appear promising. This may further enhance the numerous advantages of the spray drying process.

6.4 FGD System Reliability

Historically, FGD systems have been characterized by relatively low reliability compared with other power generation equipment. However, the technology

has evolved rapidly in the past decade and new systems going into operation are expected to demonstrate much higher reliability. This section examines the factors that have contributed to low FGD system reliability in the past and identifies measures that can be taken to improve performance.

6.4.1 Sources of Problems. The U.S. National Electric Reliability Council (NERC) has been collecting failure data on FGD systems since 1978. Table 30 identifies the largest contributing factors to plant outages for units of 400 MW and above (35).

TABLE 30 PERCENT OF ALL PLANT OUTAGE TIME DIRECTLY CAUSED BY FGD COMPONENTS (35)

Component	% Outage Time Caused
Dampers	28
Duct system, baffles	19
Fans, ID, FD, fan blades and vibration	17
Absorber tower	16
Demister	9
Reactor tank	4
Spray Pumps	3
External causes	?
Thickener	1
Test Program	1
Total	100

Problems with dampers, ducts, and baffles have been responsible for nearly half of all forced outages. Scrubber outlet ductwork can overheat due to reheat control problems, causing lining failure. When alternately exposed to wet flue gas, a highly corrosive environment is created, giving rise to materials failures. Unlike other scrubber components, redundant ductwork is typically not provided. Thus forced outages as a result of ductwork failure may be more common than with other, equally troublesome equipment.

In reviewing the U.S. EPA Utility FGD survey (9) it is apparent that corrosion, erosion, and scaling of materials are the most commonly encountered operating problems. Erosion of slurry pumps, piping and nozzles, scaling and plugging of scrubber internals and

mist eliminators, and corrosion throughout the system, but particularly at wet/dry interfaces, have all contributed to numerous failures. These have been caused by both process chemistry and materials selection problems.

During the site visits conducted as part of this study, station engineering and operating personnel also cited corrosion and erosion problems as the most common causes of system failure. However, several other problems were identified as outlined below:

1) Equipment Freezing - In some older facilities, equipment was inadequately winterized.
2) Instrumentation - Numerous facilities reported difficulties achieving reliable operation of pH meters, density meters, and SO_2 analyzers.
30 Scaling and Plugging - In several instances packed and tray-type absorbers had been converted to simple spray towers to overcome scaling and plugging of scrubber internals. Scaling and plugging of mist eliminators were also frequently identified problems.
4) Sludge Dewatering - Some facilities producing predominantly sulphite sludges and another producing a sulphate/fly ash sludge mixture experienced difficulties mechanically dewatering the material sufficiently for transport by conventional solids handling equipment.
5) Atomization - Some rotary atomizers in spray dryers encountered drive system failures and occasional poor atomization of scrubbing liquor, leading to incomplete drying and ash deposition within the absorber.

These problems have been the subject of much discussion and the object of considerable investigation on the part of the utility companies, their engineers, and equipment suppliers. While not all have been completely resolved, significant advancements have been made in the design of new systems towards alleviating many of them. Some design considerations that others have found useful in improving FGD system reliability follow.

6.4.2 Improving FGD System Reliability. Improvements in FGD system reliability are generally realized only at the expense of a higher first-cost for the system. However, this is often offset over the life of the facility by lower maintenance costs and reduced costs for replacement power, not to mention the effect on operating staff morale.

Simplicity of design and spare system capacity are both of foremost importance in achieving reliable operation. The likelihood of component failures increases with each piece of equipment added to the system. Absorber designs, for instance, have tended

to evolve towards simple spray towers with few internal components as a result of scaling and plugging problems with earlier packed and tray type towers. The spray drying process also promises to ultimately offer a highly reliable system by virtue of its very few mechanical components and the lack of wetted surfaces. However, more experience with full scale systems will be required to optimize their performance.

No guidelines are available to determine exactly how much redundant equipment should be provided in a system. However, the NERC study cited earlier suggests that the sparing concept should be applied to many more components than is currently common practice (35). Provision of a spare scrubber module has become widely accepted in the U.S. to enable an entire unit to be removed from service for maintenance without disrupting FGD operations. Sparing of slurry pumps is considered essential due to the high maintenance requirements associated with such abrasive service. Provision of completely redundant slurry loops has even been recommended to reduce the need for failure-prone block and bypass valves (35). Spare equipment obviously adds significantly to the capital cost of a system and warrants detailed analysis. Factors such as station load factor, bypass capability, replacement power cost, and predicted component failure frequency all strongly influence the selection of an appropriate sparing philosophy.

The use of forced oxidation is becoming increasingly popular as a means to both control scale formation and alleviate sludge dewatering and disposal problems. However, these are gained at the cost of higher dissolved solids and chloride content in the scrubbing liquor, which aggravates corrosion problems and may necessitate open-loop operation.

Several of the facilities visited in this study were in the northern and midwestern U.S. and experienced reasonably severe winter climatic conditions. Many station engineering and operating personnel contacted at these facilities considered complete housing of the FGD system essential to achieving reliable operation because of the difficulty of providing the high degree of equipment maintenance required under inclement conditions.

No consensus of opinion exists as to the most suitable materials of construction for use in each specific application. However, the following are some successes and failures that have been reported for specific components (35):

1) Venturi Prescrubbers - Venturi throats are usually subject to erosion, and organic coatings have generally failed. Ceramics have reportedly been more successful.
2) Absorbers - Unlined vessels of type 316 L or 317 L have been used and have the maintenance advantage of being readily repaired by pad welding. However, they are

subject to pitting and wear, particularly in low pH, high chloride environments. Carbon steel vessels with protective coatings can withstand high chloride environments, but have encountered failures due to debonding and erosion and are more difficult to repair.

3) Nozzles - These components are subject to high erosion. Silicon carbide and aluminum oxide nozzles are reportedly becoming more common, although type 316 L stainless steel has also performed satisfactorily in some instances.

4) Outlet Ductwork - Failures of organic linings have been common due to overheating and debonding. Some utilities are using high alloy metals.

5) Slurry Pumps, Piping, and Valves - Rubber-lined slurry pumps and piping have likely provided best service, although FRP piping and some alloys have also been used successfully. Few control valves have provided acceptable slurry service, although some facilities have reported most success with pinch-type valves. Wide radius bends in slurry piping are recommended to reduce wear; dead zones in which solids can settle and plug should be avoided.

Even with careful design and close attention to materials of construction, operating problems with FGD systems are inevitable, particularly in the early stages of operation of a new facility. Power plant operating personnel often lack familiarity with much of the equipment process concepts that are more typical of the chemical process industry than the power generation industry. Adequate training and a commitment on the part of management, engineering, operating, and maintenance personnel to making the system work are essential to the achievement of high FGD system reliability.

7 WASTE MANAGEMENT CONSIDERATIONS

7.1 Introduction

Chapter 5 reviewed the characteristics of some solid and liquid FGD wastes collected during site visits to facilities in the U.S. Material balances presented in Chapter 6 illustrated the quantities of these wastes that might be produced for FGD systems applied in specific Canadian situations. This chapter reviews:

a) the important factors affecting the physical and chemical characteristics of FGD wastes;

b) current waste processing and disposal alternatives applicable in Canada; and

c) the potential for waste utilization in Canada.

7.2 FGD Waste Characteristics

7.2.1 Solid Waste. The characteristics of sludges produced by the lime, limestone, alkaline fly ash, and dual alkali processes are similar for a given coal type. The sludges consist of calcium sulphate, calcium sulphite, calcium carbonate, fly ash, and trace elements. The most important factors affecting calcium-based scrubber sludge characteristics are the extent of oxidation, the amount of fly ash present, coal characteristics, reagent characteristics, and water content.

The extent of oxidation of sulphite to sulphate has a profound effect upon FGD sludge characteristics. Predominantly sulphite sludges generally have poor dewatering characteristics, exhibit thixotrophic behaviour and have little or no structural stability. Calcium sulphate sludges, on the other hand, dewater readily to a structurally stable material (gypsum) with superior physical properties.

The sulphur, ash and trace element contents of a coal also affect the composition of the FGD sludge. Changes from high to lower sulphur coals can shift the SO_3/SO_4 ratio towards the sulphate form, as excess combustion air results in proportionately more oxygen being available for oxidation. Trace elements in the coal which are partitioned into the flue gas or fly ash may ultimately be removed in the scrubber sludge.

The quantity and nature of fly ash present in scrubber sludge has a significant effect on its composition. The quantity of ash in FGD sludge decreases with increasing efficiency of upstream particulate removal devices. Larger concentrations of alkaline fly ash in scrubber sludge have been observed to impart increasing compressive strength to the sludge/ash mixture. Increased trace element concentrations in the disposed material

are also primarily a result of the inclusion of larger quantities of fly ash. Some have attributed the high sulphate/sulphite ratio, characteristic of sludges from the alkaline fly ash scrubbing process, to a catalysis effect on the oxidation reaction caused by some trace elements contributed by the fly ash.

The sorbent used in the FGD scrubber will affect sludge properties. Insoluble components in the reagent such as silica, iron oxide and alumina will appear in the sludge, as will the various trace elements which it contains. Excess calcium that does not form sulphur compounds will normally react with CO_2 in the flue gas to form calcium carbonate. Limestone, being generally less reactive than lime and thus requiring higher stoichiometric ratios, will contribute more calcium carbonate to the sludge. In the dual-alkali process, residual sodium salts added to the process in the form of soda ash reagent cannot be completely washed from the waste sludge filter cake and are retained in the disposed material.

Both the quantity and quality of water in a sludge affect its characteristics. At the optimum moisture content, a sludge generally exhibits maximum dry density and compressive strength. As the water content increases beyond the optimum, the physical properties of the material deteriorate. The quality of water in a sludge directly affects the water pollution potential of the disposed material. Dissolved species in the occluded scrubber liquor can be displaced as seepage or runoff from sludge disposal areas. The principal sources of the dissolved species found in the scrubber liquor are:

- soluble elements absorbed directly from the flue gas;
- species dissolved from the reagent;
- minor and trace elements dissolved from any fly ash present in the scrubber; and
- dissolved solids present in the makeup water.

Whereas solid wastes from the lime, limestone, alkaline fly ash, and dual alkali processes contain primarily calcium sulphate, calcium sulphite, calcium carbonate and fly ash, the purge solids from the Wellman-Lord process consist of a sodium sulphate/sodium sulphite cake. The ratio of sulphate to sulphite depends upon the oxidation rate in the scrubber and the design of the purge treatment facilities. As fly ash is removed in the prescrubber, the purge solids from the absorber loop typically contain very little ash. The purge solids are generally dried and contain very little moisture.

Solid waste from the spray drying process is produced as a dry product. When scrubbing low sulphur western coals, fly ash comprises the major component of the waste. If lime reagent is used, calcium sulphate and sulphite reaction products will be present, as

will unreacted sorbent (as hydrated lime). If soda ash is used, sodium sulphate, sulphite, and residual sodium carbonate will be present. Moisture content of the material typically ranges from 5 to 20 percent, as disposed. Relatively little oxidation of sulphite to sulphate apparently occurs in the spray dryer. For lime-based systems, the reaction products reportedly average 30 to 40 percent calcium sulphate and 60 to 70 percent sulphite.

7.2.2 Wastewaters. As described in Chapter 6, many lime, limestone, alkaline fly ash, and dual alkali wet scrubbing processes are designed to operate in a closed loop with respect to the water balance. That is, they are not designed for continuous discharge of process wastewater. Nonetheless, transient conditions may upset the water balance and necessitate a purge or blowdown. As described in Chapter 6, forced oxidation may also require a liquid purge for corrosion control. The composition of this wastewater stream would be similar to the thickener overflow or pond return water. It can be highly concentrated with respect to dissolved solids, particularly sulphite and sulphate. Appreciable concentrations of sodium, calcium, and chloride ions as well as trace concentrations of numerous elements may also be present, depending upon the makeup water quality, absorbent used, coal composition, concentration of fly ash in the scrubbing liquor, operating pH, and cycles of concentration in the scrubbing loop.

Some processes, such as the Wellman-Lord process, operate with a continuous blowdown from a prescrubber loop for the removal of chlorides and fly ash from the system. Since some SO_2 absorption occurs within this loop, it is usually operated at low pH. Although not well documented, it would appear that there is a significant potential for these low pH conditions to leach many trace metals from the fly ash into the scrubbing liquor.

The other wastewater stream that can be generated by FGD systems is leachate from the disposal of scrubber sludge or solid waste. Leachate is the liquid that has passed through or emerged from a solid waste. It may originate as precipitation or groundwater that has infiltrated the disposal site, or simply occluded water that has been released from the sludge following disposal. Leachate from laboratory tests with unfixed FGD sludge has exhibited total dissolved solids concentrations approaching those in the scrubber liquor. As the interstitial liquor is displaced by infiltrating water, the leachate becomes more dilute until the total dissolved solids concentrations have dropped to a stable level, controlled by the solubility of the major species. Laboratory tests have indicated that approximately five pore volume displacements are required to achieve a

relatively constant concentration of the major ions in the leachates (26). Because of the much higher solubility of sodium salts, compared to calcium, leachates from sodium-based FGD system wastes (e.g., sodium carbonate spray drying and Wellman-Lord) exhibit much higher dissolved solids concentrations than those from lime and limestone based systems. This observation is supported by the results of the analytical work conducted as part of this study, described in Chapter 5.

7.3 Waste Processing and Disposal Alternatives

Several environmental concerns are associated with the disposal of FGD wastes. Since solid wastes can contain certain trace toxic elements, such as cadmium, chromium, mercury, zinc, arsenic, lead, and selenium, their disposal presents a potential pollution hazard to ground and surface waters. Wastewaters generated within the scrubbing process and leachates from the solid waste disposal site can contain numerous trace elements, high dissolved solids concentrations and chemical oxygen demand which have an adverse environmental impact on the streams to which they are discharged. Another concern stems from the large quantities of waste generated and the extensive land areas required for its disposal. The poor physical characteristics of some sludges can preclude reclamation of the disposal site, meaning large land areas are covered in material with a consistency that is frequently described as resembling toothpaste. These areas are not suitable for future recreational, agricultural, or construction purposes.

A large number of alternatives are available for treating and disposing of FGD wastes. Figure 28 illustrates schematically alternatives that should be evaluated in developing a "cradle-to-grave" waste management strategy for FGD sludge. This section reviews the principal alternatives available and considered applicable to the Canadian situation for treatment and disposal of FGD solid wastes and wastewaters.

7.3.1 Solid Waste. Forced oxidation was described in Chapter 6 as an FGD process design alternative that affects the quantity and characteristics of sludge produced in calcium-based scrubbers. The process involves the injection of air into the scrubbing liquor to oxidize calcium sulphite to sulphate (gypsum), resulting in a sludge that will dewater to an 80 to 90 percent solids filter cake with vacuum filtration (26). The advantages cited for forced oxidation include not only better settling and filtering properties, but lower disposal site area requirements, good structural properties of the disposed material, minimal chemical oxygen demand in sludge leachates, and the potential for gypsum utilization as a by-product (26).

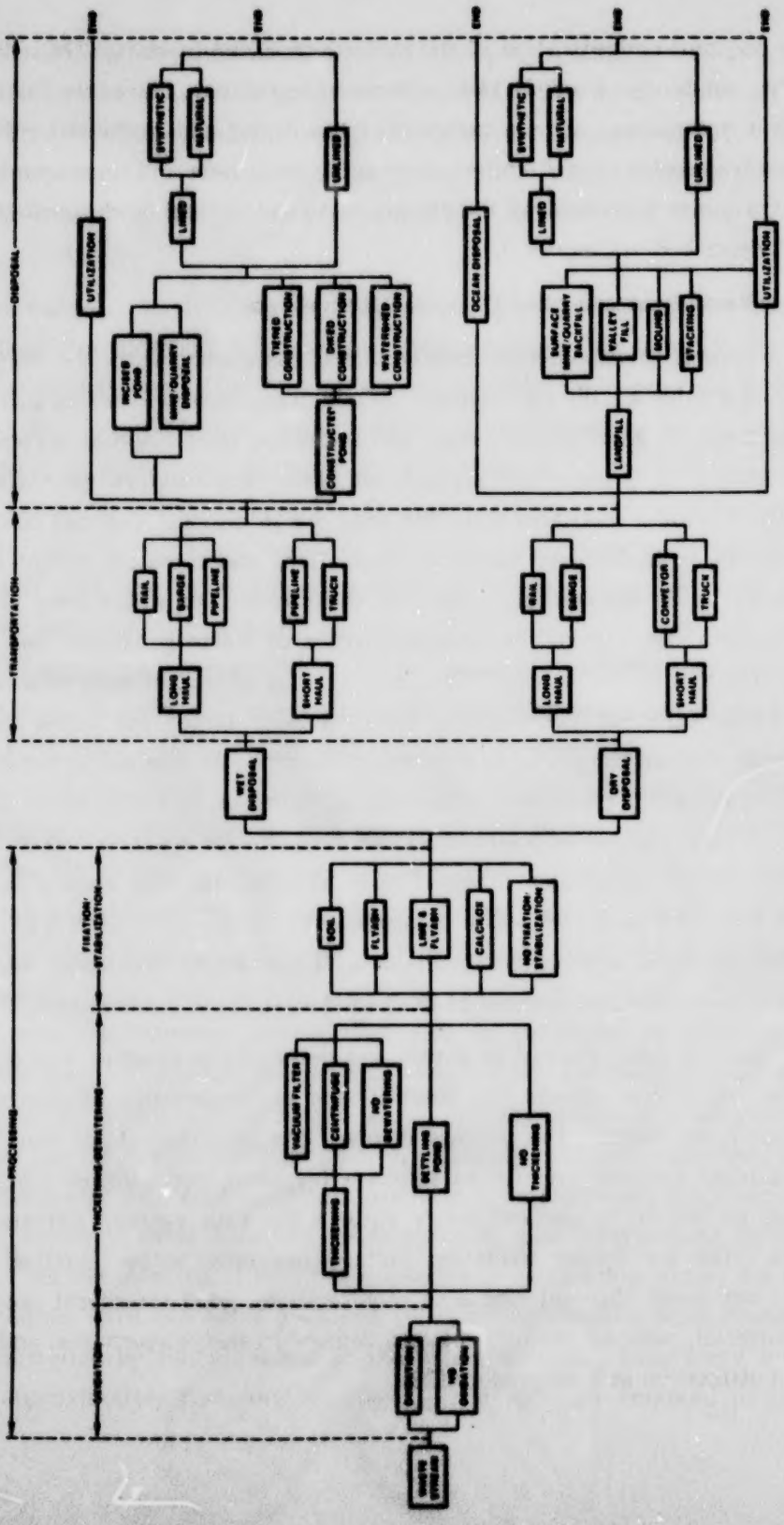


FIGURE 28 FGD SLUDGE DISPOSAL ALTERNATIVES (26)

The improved settling characteristics of gypsum crystals may permit the use of smaller thickeners than would be required without forced oxidation. Settling in ponds is also improved (settled solids concentrations in excess of 60 percent have been reported). Oxidized sludge with low concentrations of fly ash has been dewatered in ponds with underdrains to 80 percent solids in one day, yielding a material with a bearing strength of about 2000 kPa (26). Excellent dewatering results have likewise been achieved with vacuum filtration and centrifugation.

Dewatered gypsum handles well and is relatively stable when stacked. Some gypsum stacks have reached heights exceeding 30 m with side slopes of 1.5 to 1. Permeabilities in the 10^{-3} to 10^{-4} cm/s range have been reported for sedimented and cast gypsum, respectively (26).

Oxidizing calcium sulphite to sulphate largely eliminates concern regarding the high chemical oxygen demand of leachates. Sulphite is a strong oxygen scavenger, but the COD of potential leachates is greatly reduced when 90 percent or more of the sulphite is oxidized to sulphate.

Forced oxidation has been widely used in Japan to produce gypsum for use in wallboard and cement. The potential for gypsum utilization in Canada is discussed in Section 7.4. The potential disadvantages of forced oxidation are threefold. First, calcium sulphate has a higher solubility than the sulphite form. This can increase the dissolved solids content of leachate or runoff from landfills. Secondly, fugitive dust problems have been experienced in areas of gypsum landfills. However, when left undisturbed, gypsum develops a crust, which is reported to be quite resistant to both wind and water erosion. Finally, the improved dewatering characteristics of gypsum mean that more scrubbing liquor is returned to the FGD process. As described in Chapter 6, this makes closed-loop operation difficult due to the limitation it imposes on the amount of fresh water that can be added to the system and the significant increase in scrubber liquor TDS which results, giving rise to corrosion problems.

Sludge dewatering is carried out to:

a) reduce sludge volume to lower sludge transport and disposal costs as well as land area requirements;
b) permit the sludge to be handled and disposed of as a solid, rather than a liquid;
c) reduce the potential for release of sludge leachates from the disposed material;
d) recover the sludge for utilization.

Figure 29 illustrates the substantial reduction in sludge volume that can be achieved through increasing the solids content. The degree of dewatering required depends upon the type of disposal to be used. Whereas a scrubber liquor bleed stream may be sent to ponds directly or following gravity thickening, dry disposal usually requires thickening followed by mechanical dewatering.

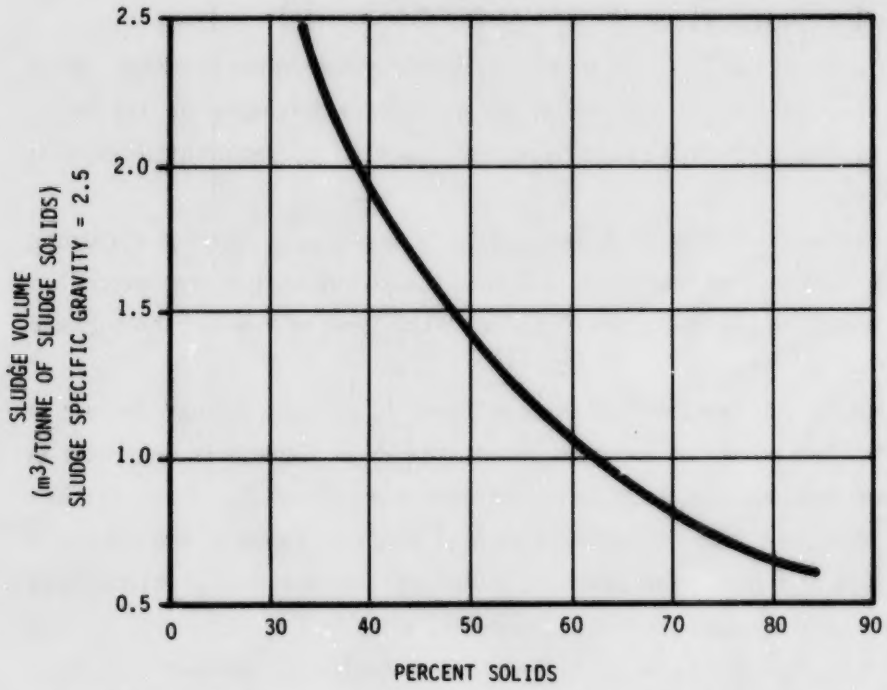


FIGURE 29 RELATIONSHIP BETWEEN SLUDGE VOLUME AND SOLIDS CONTENT (26)

Primary dewatering of scrubber bleed streams normally consists of sedimentation in gravity thickeners or ponds, or centrifugal separation in a liquid cyclone or hydroclone. While ponding may be the simplest alternative operationally, it requires a substantial commitment of land; sludge removal and disposal site reclamation can be difficult and costly; and lining for this larger area may be required for groundwater protection, resulting in high capital cost. Conventional gravity thickeners are the most common primary dewatering device. They require only a fraction of the area of settling ponds and withdraw sludge continuously. Hydroclones have been used in some new installations for solid/liquid separation. Their primary advantage is that they are less space and cost intensive than gravity thickeners. Their limitations are a higher liquid

pressure drop and the inability to separate particles less than 5 μm in size from the liquid stream (26). They are becoming more popular with forced oxidation systems for the separation of gypsum crystals with good settling characteristics.

Vacuum filters and centrifuges have both been used sucessfully for secondary dewatering. Equipment limitations that may affect the choice of one process over the other include filter cloth blinding for vacuum filters and recycle of centrate with high suspended solids concentrations from centrifuges. Both of these problems are aggravated by a high percentage of small particles in the feed stream. Capital costs generally favour the filter, with centrifuge costs being approximately 30 percent higher. Power costs are also higher for the centrifuge, but labour requirements may be higher for the filter (36).

Table 31 summarizes typical ranges of solids concentrations than can be expected from dewatering operations applied to sludges from the various calcium-based scrubbing liquors. Higher cake solids concentrations can normally be achieved in dewatering limestone sludges than lime system sludges. This is attributed to the superior dewatering characteristics of the platelet crystal structure of sulphite sludges from limestone scrubbing, compared to the rosette structure which is prevalent in lime treated sludges (26).

TABLE 31 TYPICAL SOLIDS CONCENTRATIONS OF FGD SLUDGES (26, 36)

	Lime System Sludge	Limestone System Sludge	Gypsum Sludge (Forced Oxidation)	Dual Alkali System Sludge
Scrubber Bleed	5 to 15%	5 to 15%	5 to 15%	5 to 15%
Gravity Thickening	20 to 40%	25 to 45%	60 to 65%	37 to 44%
Vacuum Filtration	34 to 62%	60 to 75%	80 to 90%	55 to 58%
Centrifuge	38 to 66%	64 to 79%	-	50 to 62%

Several approaches have been used to improve the physical and chemical properties of waste scrubber sludges to reduce the environmental impact associated with their disposal and/or to produce a by-product with reuse potential. The term stabilization has been adopted in this study to mean any treatment that induces physical changes, without chemical reaction. Fixation, on the other hand, is used to describe the addition of reagents that cause chemical reactions in the sludge.

Physical stabilization involves mixing dewatered FGD sludge with non-alkaline fly ash (i.e., low CaO content) or soil to optimize its cohesiveness and load bearing properties. As no chemical reaction is involved, the process is reversible and the material will become a fluid if rewetted. This type of sludge processing is most suitable for disposal in dry climates. Although the stabilized sludge is somewhat less permeable than untreated sludge, it is generally more permeable than chemically fixed sludge.

Chemical fixation involves mixing FGD sludge with alkaline fly ash, fly ash and lime, or a proprietary chemical. Mixing of these materials causes a cementious react on which physically encapsulates or chemically bonds the solid products into a dense and relatively impermeable matrix. Western coals typically produce an alkaline fly ash, and tests have demonstrated a direct relationship between the CaO content of the fly ash and the unconfined compressive strength of FGD sludge/fly ash mixtures. If the fly ash is deficient in calcium, significant increases in compressive strength can be achieved by adding lime to the fly ash/sludge mixture (37). Conversion Systems Inc. (CSI - formerly IUCS) offer a "Poz-O-Tec" process which uses lime and fly ash to fix FGD sludges. The Dravo Lime Company uses granulated blast furnace slag derivative, referred to as "Calcilox", to fix FGD sludges with or without fly ash.

The use of alkaline fly ash or fly ash and lime to fix FGD sludge, necessitates dewatering of the scrubber sludge prior to mixing with these materials. The optimum mix ratio required to give best physical characteristics varies depending upon the composition of the materials. Fly ash to sludge mixtures varying from 0.5:1 to 1:1 on a dry weight basis are reported typical. Lime may be added at up to 3 percent of the dry weight of the mixture (26). The opt um moisture content is one that will give the sludge its highest compacted density and normally varies from 15 percent to 40 percent after fixation (37).

The permeability of the fixed material has been found to decrease exponentially with linear increases in unconfined compressive strength as illustrated in Figure 30 (37). The physical properties of freshly placed mixtures improve with curing, and laboratory tests have demonstrated that permeabilities of 10^{-5} to 10^{-7} cm/s can be attained following 30 days curing (38). Some reports have indicated that, due to difficulties in achieving adequate compaction, field permeabilities of fixed sludge samples have been found to be somewhat higher than laboratory test results (38, 39). Compressive strengths for alkaline fly ash/sludge and lime/fly ash/sludge mixtures have been found to vary considerably, depending upon the type and amount of fly ash used, quantity of lime added, moisture content and curing time. Laboratory tests have indicated relatively low unconfined compressive strengths (i.e., less than 350 kPa) when low calcium content fly

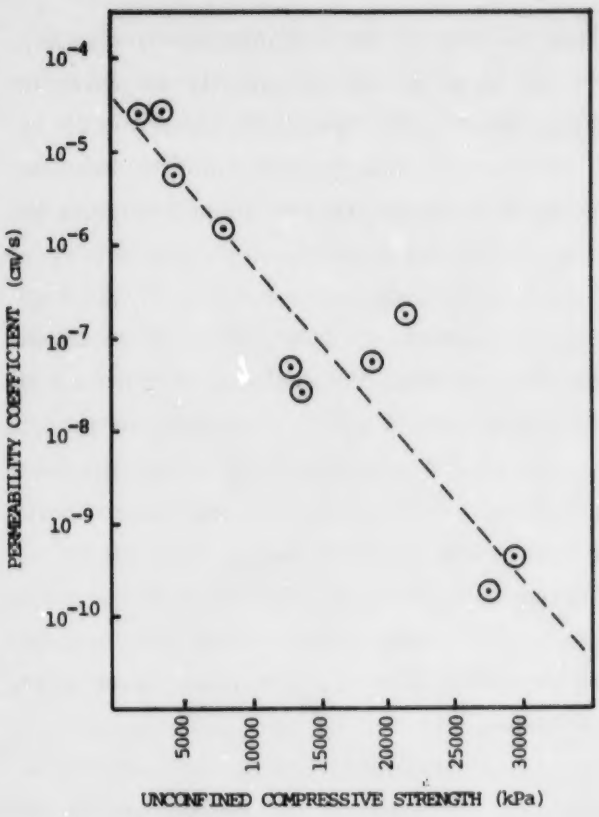


FIGURE 30 RELATIONSHIP BETWEEN FIXED FGD SLUDGE PERMEABILITY AND UNCONFINED COMPRESSIVE STRENGTH (37)

ashes derived from eastern bituminous coals were mixed with FGD sludges, whereas values in excess of 27 500 kPa were observed when alkaline fly ash from lignite and subbituminous coals was used. However, lime addition contributed to a significant increase in compressive strength for the low-calcium fly ash/sludge mixtures. In general, compressive strength increases with increasing fly ash content, increasing calcium content, and decreasing moisture content. The sulphite/sulphate ratio does not appear to influence the strength of the mixture (37).

"Calcilox" fixation can be used in either a wet or dry disposal operation. In a wet operation, Calcilox is normally added to thickener underflow at the rate of 5 to 10 percent by weight of dry sludge solids (26). The slurry, typically at 30 to 40 percent solids, is then pumped to a disposal pond, where curing occurs under water. The pond supernatant is then recycled to the scrubber or discharged. Deposited sludge reportedly achieves an unconfined compressive strength of approximately 430 kPa within 30 days

(26). Fly ash may or may not be added to the slurry in the wet operation. For dry handling, the FGD sludge is dewatered and mixed with Calcilox and fly ash prior to landfilling. In both wet and dry cases, lime addition may be required to adjust the pH to about 11, the optimum for sludge curing. Fixation of FGD sludge with Calcilox added at the rate of 10 percent by weight of dry sludge solids typically reduces the permeability by one order of magnitude, to values in the 10^{-5} to 10^{-6} cm/s range (26).

Fixation is generally considered an effective means to reduce the pollution potential of FGD sludge and make its disposal more environmentally acceptable. Improved leachate quality would be expected since some of the sludge constituents and pore water are physically encapsulated or chemically bonded in the cementious matrix. It has been reported that this reduces the solubility of the major species in the sludge and thus, the total dissolved solids concentration in the leachate, to about half that of raw sludge (26). However, the ability of fixation to improve leachate quality with respect to trace elements has been questioned. In fact, some studies have indicated that the total soluble levels of cadmium, chromium, copper, lead, mercury, selenium and zinc increase in aged FGD wastes following fixation. It is thought that some of the trace elements in the leachate may be due to the fixation additives (26, 28).

The substantial improvement in physical properties, particularly structural stability, which results from sludge fixation means the material can be disposed of by landfill. The landfill, if properly designed, further reduces the quantity of leachate generated, compared to wet disposal methods. Landfilling also facilitates reclamation and future development of the disposal area.

Fixation of waste products from dry scrubbing systems is also possible. If water is added to the waste, hydration reactions occur with unreacted reagent and calcium oxide from the fly ash. These reactions are essentially the same as those that occur in the fixation of wet scrubber sludge. The disposal-related characteristics of dry scrubbing waste products from nine test facilities were examined in one study in which samples were mixed with water and compacted (40). It was found that the optimum moisture content for the development of maximum compressive strength ranged from 10 percent to 35 percent. All samples developed a 28-day unconfined compressive strength of at least 35 kPa at some level of water addition, and values as high as 11 000 kPa were observed. In another study, 28-day unconfined compressive strengths of 80 to 12 300 kPa were obtained with lime-based spray dryer waste at moisture contents of 28 to 40 percent (41). Both studies concluded that the material developed sufficient strength to permit landfill disposal. For comparison, a person walking exerts approximately 35 kPa, whereas

a truck with tires inflated to 300 kPa will be supported by material with an unconfined compressive strength of about 200 kPa (42).

Permeabilities of wetted and cured dry scrubbing waste ranging from 10^{-4} to less than 10^{-7} cm/s were observed with most samples exhibiting values in the 10^{-5} to 10^{-6} cm/s range (40, 41). Permeability coefficients of 10^{-7} cm/s are required by the U.S. EPA for liner materials in nonhazardous waste impoundments (40).

When cured and uncured samples were subjected to the U.S. EPA extraction procedure, the extracts of cured waste products usually had lower concentrations of most elements than those of untreated material (40). Although the study did not specifically address the differences between extracts of samples from lime-based systems and those from sodium carbonate systems, significantly higher concentrations of TDS would be anticipated in the latter, due to the much higher solubility of sodium salts.

Final disposal of FGD waste solids may be either a wet or a dry operation. In wet disposal systems, the solids are transported to and deposited in a pond as a slurry. The scrubber bleed stream may be sent directly to the disposal pond, or it may be prethickened to reduce the pond size. Alternatively a slurry of thickened sludge and fixation chemicals may be sent to the pond for disposal. Solids are allowed to settle in the pond and supernatant is normally recycled to the FGD process. Ponding is applicable to lime, limestone, and alkaline fly ash wet scrubbing systems. However, dual alkali systems are not normally suited to wet disposal since the sludge must be dewatered and washed to recover the scrubbing reagent.

The principal advantage of ponding is the simplicity of a totally pumped disposal system compared to a dewatering and materials handling system. Variations in input stream flow or composition can be attenuated, and because ponds require minimal operational attention and maintenance, operating costs are relatively low. The capital cost of the disposal pond may, however, be substantial. Large land areas are required, construction involves much earthwork, and an impervious liner may be required to contain leachate if native soils are unsuitable.

The problems of maintaining the scrubber water balance with a pond system in a wet climate were addressed in Chapter 6. In addition, the hydraulic head of standing water in the disposal pond creates the potential for the generation of large volumes of leachate. A pond liner may be necessary to protect the underlying groundwater regime. Some utilities have also reported scaling in pond water return lines.

Reclamation of many disposal ponds is likely to be difficult if not impossible. Sludges with a high sulphite to sulphate ratio are thought to remain perpetually

thixotropic and structurally unstable. This may effectively remove large areas of land from any future use and pose the ongoing problem of a perpetual discharge from the retired disposal basin.

Wet disposal with chemical fixation (e.g., "Calcilox") may help alleviate some of the concerns expressed above. It can reduce leachate generation by decreasing the permeability of the deposited material and can increase its structural stability to facilitate reclamation. Chemical fixation should be considered for any wet disposal system, but particularly where a high sulphite/sulphate ratio is anticipated in the FGD sludge.

One design option that has been used to facilitate sludge dewatering and pond reclamation is the installation of an underdrain system. This enables free water occluded in the sludge to drain through the accumulated material and be removed upon retirement of the pond. Such a system is illustrated in Figure 31. This may be particularly effective with alkaline fly ash processes which typically have highly oxidized sludge with reasonably good drainage properties.

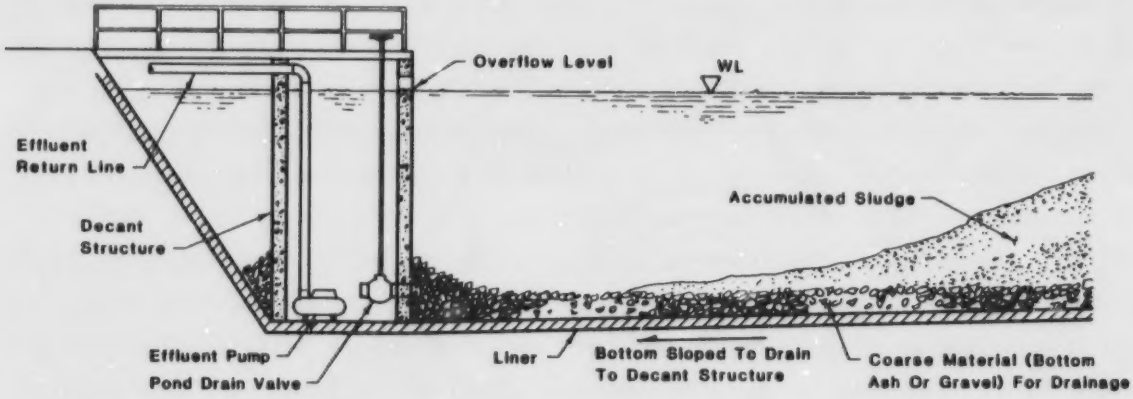


FIGURE 31 POND UNDERDRAIN AND DECANT STRUCTURE

Dry disposal involves the transport of FGD waste as a solid to a disposal site, where it is placed and compacted in a landfill. Landfilling is only applicable to structurally stable materials such as dewatered sulphate sludges, sulphite sludges that have undergone stabilization or fixation, and waste solids from dry scrubbing processes. Landfills require significantly less land area than do ponds because the waste material is denser and retaining structures such as dams and dikes are not necessary. They can generally be reclaimed readily by covering with loam and revegetating.

The potential for leachate generation is substantially reduced in landfills, since there is no head of standing water above the wastes and the permeability of the compacted material is generally lower than that of ponded material. If the landfill is developed and reclaimed in smaller cells or modules, the amount of material exposed to the effects of precipitation at any given time is relatively small. Designing the facility to divert runoff away from the landfill site further reduces leachate potential. Nonetheless, impervious liners may be required to attenuate the migration of pollutants into the underlying groundwater. A permeability of 10^{-7} cm/s should normally provide the required protection (26). The layout and design of a landfill disposal site is illustrated in Figure 32.

The principal disadvantages of dry disposal systems are their relatively high operating costs, and the noise, dust, and aesthetic impact of landfill operations. Furthermore, not all sludges are amenable to disposal by landfill. Predominantly sulphite sludges lack the structural integrity to support a landfill operation and material handling problems have even been experienced with some sulphate sludges. The Tennesee Valley Authority is experimenting with wet and dry stacking of sludge from a forced oxidation system. The material contains about 25 to 30 percent fly ash and initial results with mechanical dewatering and landfill have been discouraging, particularly in wet weather (43). Vacuum filtration will dewater the material to about 80 percent solids. However, when placed in a temporary stacking area, it has regained up to 40 percent moisture from precipitation, rendering it difficult to handle and virtually impossible to compact. A parallel wet stacking test is also being carried out, but its relative success had not yet been evaluated at the time this report was written (43).

With the advent of forced oxidation FGD systems, stacking methods for gypsum disposal which have been employed in the phosphate fertilizer industry have been considered. This disposal technique involves pumping the waste gypsum slurry to a sedimentation pond, where the solids are allowed to settle and drain, and process water is decanted and returned to the scrubber. When sufficient gypsum has been deposited, a drag line is used to excavate the settled solids, which are then used to raise the perimeter dike. Stack heights in excess of 30 m with side slopes of 1.5:1 have been reported (26). Using this approach, the material is essentially stockpiled for possible future use. A typical design and construction technique for a gypsum stack is illustrated in Figure 33.

As gypsum is relatively permeable, water seepage through the stack can be significant. To prevent groundwater contamination beneath the stack, it should either be constructed on low permeability soils or a liner should be installed. A seepage collection

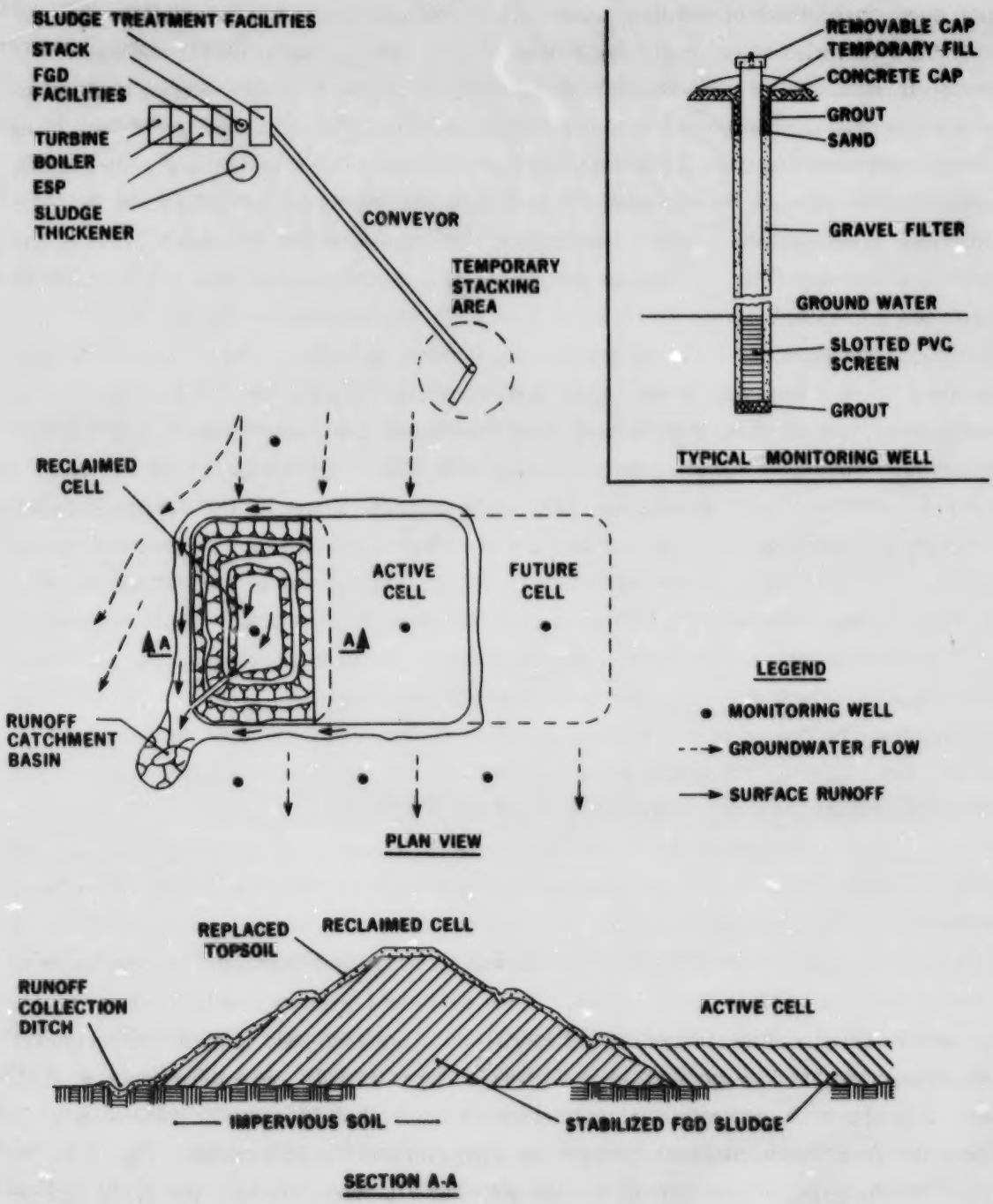
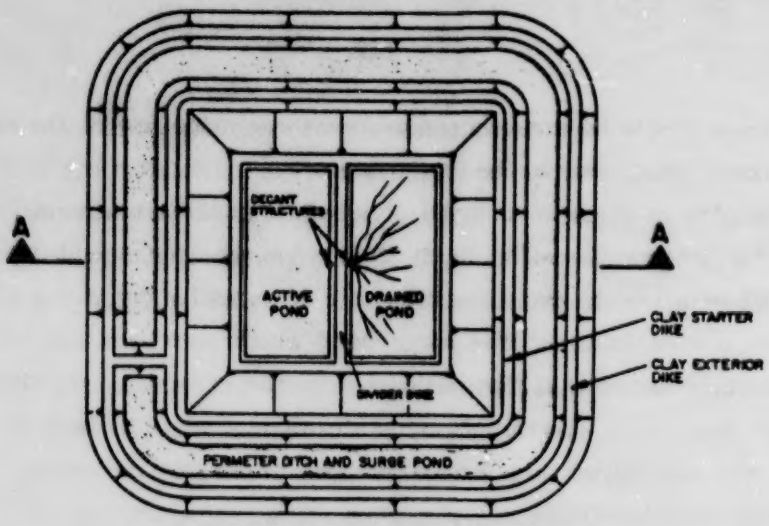
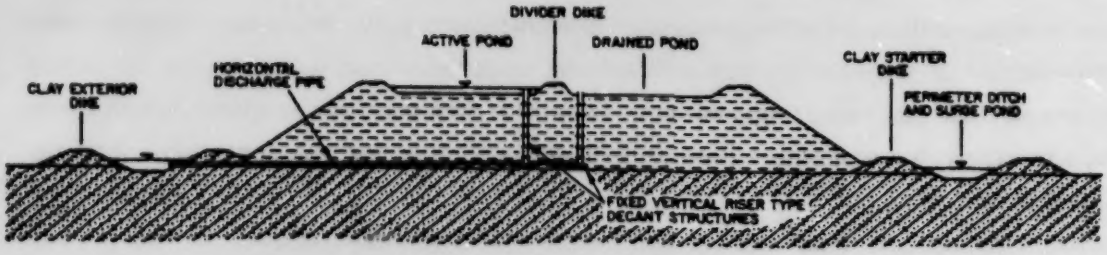


FIGURE 32 LAYOUT AND DESIGN OF A TYPICAL FGD SLUDGE LANDFILL



GYPSUM STACK SITE PLAN



SECTION A-A GYPSUM STACK CROSS SECTION

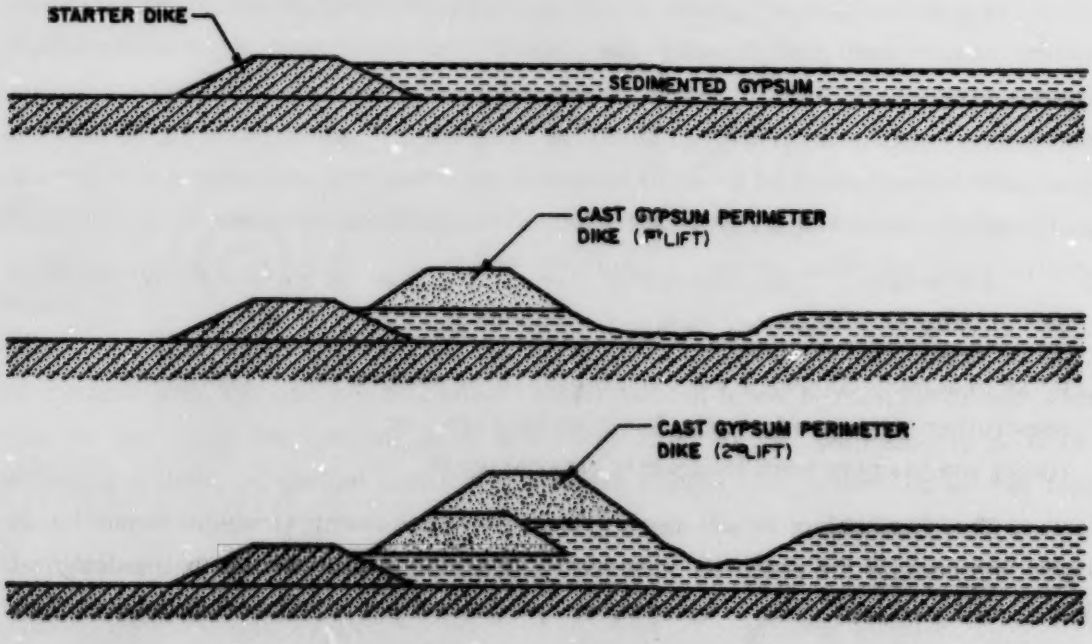


FIGURE 33 DESIGN AND CONSTRUCTION OF A TYPICAL GYPSUM STACK (26)

ditch (surge pond) should be constructed around the outside toe of the stack to collect seepage and runoff for recycle to the FGD system.

The gypsum stack is normally divided into at least two ponds, so that one can drain while the other is in use. The depth of water in the active pond is normally kept to a minimum (0.3 to 1.0 m) to avoid the discharge of excessive quantities of process water in the event of a dike failure. The surge pond which surrounds the stack is typically designed to contain the normal fluctuations in operating levels, depending on the water balance, and 1.5 times a 25-year, 24-hour storm on the watershed of the pond (44). Sufficient storage capacity should also be provided to prevent the release of water in the event of a perimeter dike failure on the gypsum stack.

A number of phosphate fertilizer plants with gypsum stacks are located throughout the northern United States and Canada. No deleterious effects on the gypsum slopes have been reported after successive freeze-thaw cycles. Sufficient capacity must be provided within the stack to store all gypsum produced during winter as the perimeter dikes are not raised in freezing weather. Greater water depths may also be required to allow a sufficient zone of water flow beneath the ice cover for sedimentation, clarification, and decanting to occur.

The effect of fly ash addition on the stacking characteristics of FGD gypsum has been investigated. Unlike the addition of fly ash to sulphite sludges, it reportedly decreases the sedimentation/consolidation and the shear strength of gypsum (39, 44).

The advantages of gypsum stacking are simplicity of operation and minimal equipment requirements, which offer low capital and operating cost as well as high reliability. The principal disadvantage, compared to dry disposal, is that precipitation inputs to the system are more difficult to avoid. All water entering the stack automatically becomes part of the FGD system water inventory, aggravating the already difficult task of maintaining a closed water loop with forced oxidation systems.

7.3.2 Wastewater. The wastewater streams that can be generated by the FGD processes considered in this study include the following:

a) purge of scrubbing liquor from lime, limestone or alkaline fly ash processes;
b) prescrubber blowdown from the Wellman-Lord process;
c) runoff and leachate from solid waste disposal areas.

The first objective of any wastewater management strategy should be to minimize and contain the potential wastewater streams. Recent trends in the design of FGD systems have been towards closed-loop systems. Considerable development effort

has been directed towards maximizing process water reuse and the selection of materials of construction that are tolerant of the high TDS concentrations characteristic of closed-loop operation. The state-of-the-art of FGD process design appears to be sufficiently advanced that closed-loop operation is technically feasible and attainable with conventional wet scrubbing systems operated in a natural oxidation mode with medium to higher sulphur coals.

Another trend which has emerged in recent years is the favouring of forced oxidation over natural oxidation systems. While this offers significant environmental advantages with respect to solid waste disposal, it increases the likelihood of generating a wastewater stream from the process. As previously discussed, the improved dewatering characteristics of gypsum sludge mean that substantially more scrubbing liquor will be recycled to the process, reducing both the quantity of fresh makeup water that can be added and the blowdown of dissolved solids leaving the system as occluded water in the sludge. To maintain a closed water loop, greater quantities of higher TDS water must be reused within the process. Considerably more experience with forced oxidation will be required to demonstrate the feasibility of operating FGD systems in this mode without a liquid purge.

The same is true for conventional (i.e., natural oxidation) systems operated with low sulphur coal. The relatively small quantities of solid waste leaving the system means that correspondingly small volumes of scrubber liquor are purged as occluded water. FGD systems applied to low sulphur fuels generally operate at several cycles of concentration higher than the same system applied to high sulphur fuel.

While transient conditions that create temporary water imbalances cannot be avoided, FGD processes can be designed with sufficient surge capacity to accomodate many of these conditions and enable scrubbing liquor to be recovered rather than purged. An emergency dump pond is advisable to contain scrubbing liquor from a temporary water imbalance, major equipment failure, or draining of process tanks for maintenance or repair.

Solid waste disposal sites can be designed to minimize the generation and loss of leachate from the site, and to contain runoff. Loss of leachate from ponds and landfills can be controlled by ensuring they are constructed in low permeability soils or by installing a liner. Chemical fixation of sulphite sludges can further aid in this regard. Construction and reclamation of landfills and gypsum stacks in smaller modules or cells can further reduce leachate generation by reducing the time of exposure of disposed material to the effects of precipitation.

Ponds and gypsum stacks should be constructed with sufficient freeboard capacity that precipitation from a reasonably conservative design storm can be accomodated without overflow. Landfills should be graded to prevent uncontaminated overland flow from infiltrating the disposal site, while containing runoff from the active landfill area. This can be accomplished using ditches and ponds or containment dikes.

If wastewater flows have been minimized and contained, the choice must then be made to either reuse them or to dispose of them. In general, if the characteristics of a wastewater stream preclude its reuse, it will in all likelihood be unsuitable for discharge to the environment without prior treatment. There may be some exceptions. For instance, if the runoff from a landfill operation can be demonstrated to be non-toxic and relatively free of contaminants, it may be acceptable for direct discharge. The size of the wastewater stream and its location relative to the principal water-consuming operations of the FGD process may render reuse of the stream impractical.

An evaluation of the treatment requirements for FGD wastewaters must begin with a review of discharge criteria. Environment Canada's proposed criteria for discharge of wastewaters from new thermal power generating stations are summarized below:

pH	6.5 - 9.5
Chromium	0.5 mg/L
Copper	0.5 mg/L
Iron	1.0 mg/L
Nickel	0.5 mg/L
Zinc	0.5 mg/L
Suspended Solids	25 mg/L

Although these are the only federal guidelines that would apply to FGD wastewater discharges, other parameters which may be of environmental concern (and for which treatment may be necessary to meet other regulatory requirements) include COD, TDS, and other heavy metals. Environment Canada does not consider dilution of FGD wastewaters in condenser cooling water or other high volume, uncontaminated streams an acceptable practice for protection of the aquatic environment. These wastewaters should be treated for contaminant removal at their source.

Lime treatment followed by filtration is likely to be the most cost-effective treatment method for heavy metal removal. This has been used in Japan for the treatment of FGD system blowdown streams (29). The limited treatability studies conducted as part of the laboratory work described in Chapter 5 also demonstrated the

ability of lime treatment/filtration to reduce heavy metals to acceptable levels, judged by the above criteria. Lime treatment is usually considerably less expensive than ion exchange or other technologies for heavy metals removal from wastewater streams of this type.

The principal chemical specie contributing to COD in FGD wastewaters is normally the sulphite ion. As this is readily oxidized to sulphate by the introduction of oxygen, simple aeration may be sufficient to reduce COD to acceptable levels for discharge. In Japan, FGD wastes are reportedly treated by ion exchange for the removal of dithionate ($S_2O_6^{=}$) which contributes to COD (29). No North American facilities are known to employ this method for COD removal.

Where wastewater streams are created by a need to control dissolved solids accumulation in the FGD system, the TDS content of the liquid purge may be so high as to preclude discharge to receiving waters. TDS reduction can only be achieved through the application of relatively expensive evaporation or membrane process technology. Vapour compression evaporators have been installed in electric power generating stations in the western and southwestern United States for treatment of cooling tower blowdown. They have also been evaluated for application to the treatment of purge streams from FGD systems. One study concluded that the technology is applicable to the removal of soluble salts and trace metals from purge streams that may result from the use of forced oxidation with high chloride coals (45). A vertical tube, falling film vapour compression evaporator was also used to concentrate liquid wastes from the Chiyoda Thoroughbred 101 process at an EPRI sponsored pilot project at Gulf Power's Scholz Generating Station (46). Over 99 percent of the waste was recovered as product water with less than 10 mg/L of TDS. While treatment by evaporation is costly, the product water would likely be of sufficient quality that it could be rerouted to the boiler feedwater treatment plant. The costs associated with wastewater treatment would then be at least partially offset by reduced pretreatment costs for boiler feedwater treatment.

In general, wastewater treatment for FGD systems have been given very little attention in North America. This may be because most efforts have been directed towards developing a process design for closed-loop operation. However, with the advent of forced oxidation processes, a need exists to examine wastewater treatment more closely.

The final step in an effective wastewater management program is to monitor the quality and the effects of any discharge to the environment. For instance, if disposal site runoff is to be impounded and intermittently released, the composition and volume of

the wastewater should be determined prior to discharge. This will provide an ongoing record of the effectiveness of the disposal operation in mitigating water pollution problems. It will also identify the need for treatment of the waste prior to discharge. Groundwater monitoring is also recommended in the vicinity of the disposal site to determine the impact leachate may have on underlying aquifers. Baseline data on groundwater quality should be gathered prior to the commencement of disposal operations, and ongoing routine monitoring should be performed. The number and location of wells required to adequately monitor the groundwater regime must be determined by a site-specific hydrogeology study.

7.4 Waste Utilization

Four groups of scrubber wastes or by-products generated by the FGD processes considered in this study have reuse potential. These are:

a) elemental sulphur and sulphuric acid from the Wellman-Lord process;
b) fixed sulphite sludge;
c) dry scrubbing wastes; and
d) gypsum from forced oxidation processes.

The enormous and increasing quantity of FGD wastes and by-products produced annually in the United States has generated interest in the utilization of these materials. Such by-product utilization has been practised in Japan for several years. This section addresses some of the technical and market factors which may affect FGD waste utilization in Canada.

7.4.1 Elemental Sulphur and Sulphuric Acid. Sulphur, an important industrial raw material, occurs naturally in both elemental and combined forms. Almost two-thirds of the world's sulphur is in elemental form, nearly all obtained from native sulphur deposits or from treatment of "sour" natural gas. The remainder is recovered from pyrite and smelter stack gases, principally as sulphuric acid (H_2SO_4). Approximately 87 percent of all sulphur produced is utilized as sulphuric acid. Fertilizer manufacture accounts for over half of all sulphur consumed, with the manufacture of chemicals, pigments, and pulp and paper as the next largest consuming sectors (47).

Canadian sulphur is obtained from three sources: elemental sulphur derived from sour natural gas and petroleum; sulphur recovered from smelter gases in the form of sulphuric acid; and sulphur contained in pyrite concentrates used in sulphuric acid manufacture. Minor tonnages of elemental sulphur are recovered as a by-product of

electrolytic refining of nickel sulphide matte, and about 85 000 tonnes per year of liquid sulphur dioxide are produced from smelter gases. Canada has been the world's largest exporter of elemental sulphur since 1968 (48). In 1978, 90 percent of Canadian sulphur shipments were in elemental form, nearly all from sour natural gas in western Canada.

For the 7 percent sulphur coal (type 1), and 2 percent sulphur coal (type 4) considered in this study, a 500-MW Wellman-Lord process would produce 176 800 tonnes/annum and 39 200 tonnes/annum of sulphuric acid, respectively, assuming a 65 percent load factor. In comparison, the 1977 Canadian consumption of sulphuric acid was 2 786 447 tonnes.

It is concluded that a viable market would probably exist for by-product sulphur or sulphuric acid in some regions of Canada. However, the high cost of FGD using the Wellman-Lord process is such that even if all by-product sulphur was sold at $80/tonne (49), the process would still be significantly more expensive than the alternatives available. For the 7 percent sulphur coal case, it is estimated that by-product sulphur sales would represent a cost credit of approximately 1.6 mills/kWh, whereas for the 2 percent sulphur coal case the credit would be only 0.4 mills/kWh. As will be illustrated in Chapter 8, this is insufficient to offset the unfavourable economics of the Wellman-Lord process in most cases.

7.4.2 Fixed Sulphite Sludge. Researchers have experimented with several uses for fixed sulphite sludges in the United States (26). The relative impermeability and good structural properties of the material have led to demonstration projects using the material for paving, wastewater pond lining, concrete block construction and general fill. Although several of these applications appear technically feasible, the overall demand for sludge for these uses is not expected to approach that which could be supplied. In addition, there is seldom much economic incentive to use waste sludges in place of most construction materials. This usage of sludge is expected to be limited and will depend largely on local conditions.

7.4.3 Dry Scrubbing Wastes. The use of dry scrubbing wastes as concrete additives and synthetic aggregate has been investigated to a limited extent. The spray drying process has been applied primarily to western low sulphur coals with alkaline ash. Since the largest component of the waste material is fly ash, and the mixing of alkaline fly ash with portland cement is already practised commercially in the production of some concretes, there is good reason to expect that at least a limited market will exist for this material. During the station visits conducted as part of this study, one utility indicated

that interest had been expressed by outside companies in the use of their spray dryer waste product for oil well cementing applications, and that investigations were currently under way to investigate its suitability.

The pelletizing of dry scrubbing wastes to produce synthetic aggregate has also been investigated. By adding 10 to 20 percent moisture to the dry waste and mechanically compressing it to between 2.5 and 3.0 times its loose density, high strength pellets have been produced, suitable for use in cement mixes and road base applications (42). The cost of pellet production was estimated at $6.00 U.S./tonne (1982) based on a levelized cost over a 30-year economic life (41). This was considered competitive with the cost of crushed stone. The marketability of the material in Canada will obviously depend upon the cost and availability of conventional sources of aggregate, which will be highly site-specific.

7.4.4 Gypsum. Virtually all FGD systems associated with power plants in Japan produce commercially reusable by-products, with gypsum being the most common (50). While unique market and technical considerations exist in Japan, with respect to lack of indigenous gypsum and land availability, FGD by-product gypsum use also appears to be of increasing interest in Germany and North America.

Gypsum production has corresponded closely to activity in the building construction sector, where wallboard, the principal gypsum product, is used in both residential and non-residential buildings. The use of gypsum wallboard has also been increasing in institutional and commercial buildings because of its fire retardant properties. Gypsum is also used as a setting retardant in cement.

Recently, the gypsum-board industry has been relatively depressed due to the generally depressed economy of the construction industry. In 1981, the United States output of crude gypsum decreased by 3 percent. Of the total gypsum supply (about 20 million tonnes), 74 percent was used in prefabricated building products, 15 percent was used in cement, 7 percent was used in agriculture, and 4 percent was used in industrial and building plaster. Approximately 5 percent of the gypsum used was by-product gypsum from industrial sources (51).

Canadian gypsum production, trade and consumption figures for 1979 are presented in Table 32. The estimated gypsum production from a 500-MW gypsum by-product FGD process using 7.0 percent, 4.0 percent, 0.5 percent, and 2.0 percent sulphur fuels are also shown. This would represent between 0.3 percent and 4.0 percent of the 1979 Canadian production, and between 1.0 percent and 12.0 percent of consumption for that year.

TABLE 32 CANADIAN GYPSUM PRODUCTION, TRADE, AND CONSUMPTION, 1979 (52)

	tonnes
Production	8 105 000
Imports	152 463
Exports	5 474 736
Apparent Consumption	2 782 727

Notes: 1) Apparent Consumption = Production + Imports - Exports.
2) FGD by-product gypsum production from a hypothetical 500 MW station at a 65 percent load factor would be as follows:

Coal Type	% S	$CaSO_4 \cdot 2H_2O$ (tonnes/annum)
1	7.0	327 000
2	4.0	155 000
3	0.5	27 400
4	2.0	72 900

Because gypsum is a relatively low-cost, high-bulk mineral commodity, it is generally produced from natural deposits situated as close as possible to areas where markets exist. Exceptions occur where deposits of unusually high quality are available, even at a somewhat greater distance from markets, where comparatively easy and inexpensive mining methods are applicable, or where low-cost, high-bulk shipping facilities are accessible. Nova Scotia and Newfoundland deposits meet all three of these criteria and have been operated for many years by, and for, United States companies in preference to United States deposits. Crude gypsum, mainly from the Newfoundland port of St. George's, and from Halifax and Little Narrows in Nova Scotia, is also shipped to the Montreal and Toronto areas for use in gypsum products manufacture and portland cement production (52). Shipping distances for finished gypsum products are normally kept to a minimum because freight and handling costs are a major part of the price to the consumer for items that are relatively low-priced and readily available at many locations.

Gypsum as a by-product from FGD systems is generally similar to naturally occurring gypsum deposits. However, its physical and chemical characteristics can differ depending on the fuel used at the station, the scrubber absorbent, and the FGD system operating characteristics. The chemical characteristics of by-product gypsum must correspond to certain requirements. Constituents that negatively affect the use of by-

product gypsum include chlorides, fluorides, sodium, magnesium, calcium sulphite, and fly ash. For example, chloride adversely affects calcining temperature, set time, and stucco slurry consistency (29). Sodium and magnesium have efflorescent properties and sulphite has adverse crystal-modifying effects. These contaminants cause reduced strength and unfavourable behaviour when applying gypsum-plaster, and can also affect secondary reactions during calcining (53). The presence of more than 1-2 percent fly ash can cause wallboard discolouration (26). Efficient particulate removal ahead of the FGD system becomes an essential design component for by-product gypsum systems. FGD system additives may also adversely affect the use of by-product gypsum, particularly for wallboard manufacture. Carboxylic acid is reportedly a setting retarder, whereas valeric acid (a degradation product of adipic acid) is odourous (54).

Table 33 shows the specifications of three U.S. companies and the German Gypsum Association for reuse of FGD sludges as gypsum replacement (55, 56). None of these organizations place limits on concentrations of trace or heavy metals.

TABLE 33 SPECIFICATIONS FOR FGD BY-PRODUCT GYPSUM (55, 56)

Parameter	Georgia Pacific	National Gypsum	U.S. Gypsum	German Gypsum Association*
Minimum purity ($\%CaSO_4 \cdot 2H_2O$)	90	94	N/A	94.5
Maximum sodium (mg/kg)	200	500	75	590
Maximum chloride (mg/kg)	200	800	75	100
Maximum magnesium (mg/kg)	N/A	500	50	603
Maximum free water (%)	10	10 (Max. 15)	10 ± 2	10
pH - minimum	3.0	6.0	6.5	5.0
- maximum	9.0	8.0	8.0	9.0
Calcium Sulphite (%)	N/A	0.5	N/A	N/A
Particle size (μm)				
- minimum	N/A	N/A	20	N/A
- maximum	N/A	N/A	40	N/A

* Bundesverband Der Gipo-und Gipbauplattenindustrie E.V.
N/A - no specification available

A major obstacle to FGD by-product gypsum utilization is the high chloride content of some sludges. Chlorides originate in the coal and the makeup water (particularly if station wastewaters are used as makeup to the FGD system). The greater degree of dewatering associated with forced oxidation reduces the "blowdown" of dissolved solids occluded with the sludge from the FGD system, thereby increasing the chloride levels. It has been suggested that chloride levels of 25 000 - 100 000 mg/L could be expected in a typical eastern bituminous fuelled station with a coal sulphur content of 3 percent (57). To achieve final chloride levels below 200 to 300 mg/kg in such cases it is reported that the cake must be reslurried and rewashed, increasing the cost of sludge treatment and the volume of wastewater generated (55). Open-loop operation of gypsum by-product FGD systems has been recommended to reduce scrubbing liquor chloride concentrations to 2000 to 5000 mg/L (29). Coupled with less efficient cake washing, this may be a less expensive alternative depending upon wastewater treatment requirements. Japanese systems normally use a prescrubbing loop for chloride removal. Regardless of the approach, the low chloride requirements for by-product FGD gypsum will adversely affect the system water balance. Considerably more work is required to determine the extent of this impact and the most cost-effective means of handling the resultant wastewater.

A higher chloride content gypsum could be used as a cement setting retardant if the gypsum were not washed to a chloride level suitable for gypsum board. However, the market for this application appears to be less profitable than for gypsum board (56).

Handling of by-product gypsum may also present some problems. Existing equipment in the gypsum industry is generally designed to handle dry and lumpy natural gypsum rock. By-product gypsum is a moist, fine particled powder. Agglomeration and drying of the wet by-product to facilitate handling in some existing gypsum processing facilities may be required.

Because FGD sludge production would be greatest during periods of greatest station load, usually in winter, an intermediate storage facility for by-product gypsum would be required for the period between maximum production and consumption.

In summary, several technical and ecomomic factors must still be addressed to fully evaluate the potential for reuse of by-product gypsum. However, the prospect for waste utilization holds sufficient promise to warrant further investigation.

8 FGD SYSTEM COST

8.1 Sulphur Dioxide Removal Costs

FGD system cost has been the subject of considerable study and controversy in North America in recent years. Numerous published reports have attempted to estimate the comparative costs of alternative FGD processes. The approaches have varied from process predesign with procurement of vendor quotations for major equipment followed by detailed estimation of installation costs, to a simple scaling of actual or known costs from a similar facility to account for differences in plant size, coal type, or other site specific parameters. In 1978, Battelle Columbus Laboratories prepared a detailed analysis of variations in costs of FGD systems for the Electric Power Research Institute (58). Process factors identified as affecting cost variations included plant size, coal sulphur content, degree of sulphur removal, operating variables, project scope, sludge disposal, by-products, and retrofit problems. Some of the conclusions reached regarding the relative effects of these factors are summarized below.

1) The unit investment cost in $/kW is about 30 percent less for a 1000-MW, ten-module system than for a 200-MW, two-module system.
2) An increase in coal sulphur content from two percent to five percent increases the unit cost for a 500-MW plant by about 31 percent.
3) The cost of a scrubber system on a new 1000-MW plant fired with 3.5 percent sulphur coal and designed to meet a 90 percent sulphur dioxide removal standard will be about 5 percent greater than a similar system designed to meet a 520 ng/J standard.
4) Variation in the scope of facilities included in the FGD system cost estimate accounts for a substantial proportion of the reported variation. This may be due not only to variations in the portion of the total power generating facility included in the FGD project scope, but variation in several other important design criteria such as the degree of equipment winterizing required (e.g., insulating, heat tracing, and housing of facilities), redundancy philosophy, etc.
5) Retrofit problems may increase the cost estimate by as much as 50 percent over the cost of a new facility.
6) The method of sludge disposal selected and the location and geotechnical conditions of the disposal site contribute to variations in FGD system total cost.

7) Sulphur and sulphuric acid by-product sales can reduce operating costs by as much as 1.0 mills/kWh.

Other conditions such as bidding conditions and financing arrangements may also affect FGD system cost.

Many factors can frustrate an attempt to develop generalized conclusions regarding the relative costs of alternative sulphur dioxide control technologies from published data without detailed site-specific information. Scaling of published U.S. data to reflect Canadian conditions at a different point in time introduces yet another source of variance. In a generic study such as this, the costs presented are intended to represent relative order-of-magnitude values for comparative purposes only and should not be treated as definitive estimates.

To illustrate the relative costs of the principal FGD process alternatives considered in this study two sources of generalized cost data were used: a study prepared by the Tennessee Valley Authority (TVA) for NATO (59), and cost information reported by U.S. utility companies and compiled by PEDCo Environmental Inc. for the U.S. Environmental Protection Agency (9). The reader is referred to these publications for detailed information concerning the assumptions and approach used in preparing these estimates. However, the key elements of the cost basis used in each reference are summarized below.

The economic evaluations prepared by TVA for NATO were based on a new 500-MW power plant assumed to have a 30-year lifetime. Several FGD processes and coal types were evaluated. The costs were based on a Midwestern U.S. location using mid-1979 as the basis for capital costs and mid-1980 as the basis for all direct and indirect costs for equipment, land, materials, labour, fees, services, and other construction costs required to install the FGD system. Annual revenue requirements were based on a 6000 hour per year operation and consisted of all raw material, labour, utility, and other conversion costs, and indirect costs such as capital charges, taxes, and overheads. These costs were converted to 1983 Canadian values by adjusting for inflation (25 percent for capital costs and 17 percent for operating costs) and for U.S. to Canadian exchange (at 25 percent). The results are presented in Figures 34 and 35, which illustrate the capital costs and annual revenue requirements, respectively. The range of costs illustrated in these figures reflect TVA's predicted accuracy of the estimate for each process. In general, processes for which more operating experience and actual cost information were available exhibit a smaller range of expected accuracy than those for which little data of this nature was available.

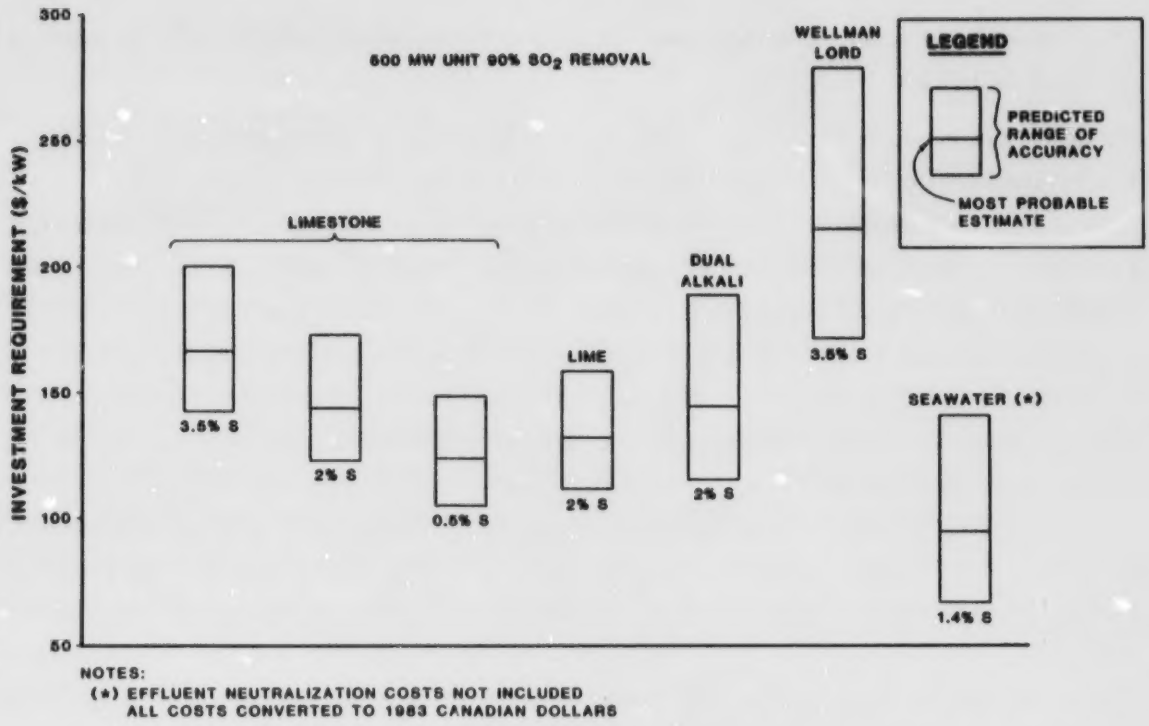


FIGURE 34 CAPITAL INVESTMENT REQUIREMENTS FOR ALTERNATIVE FGD PROCESSES REPORTED BY NATO (59)

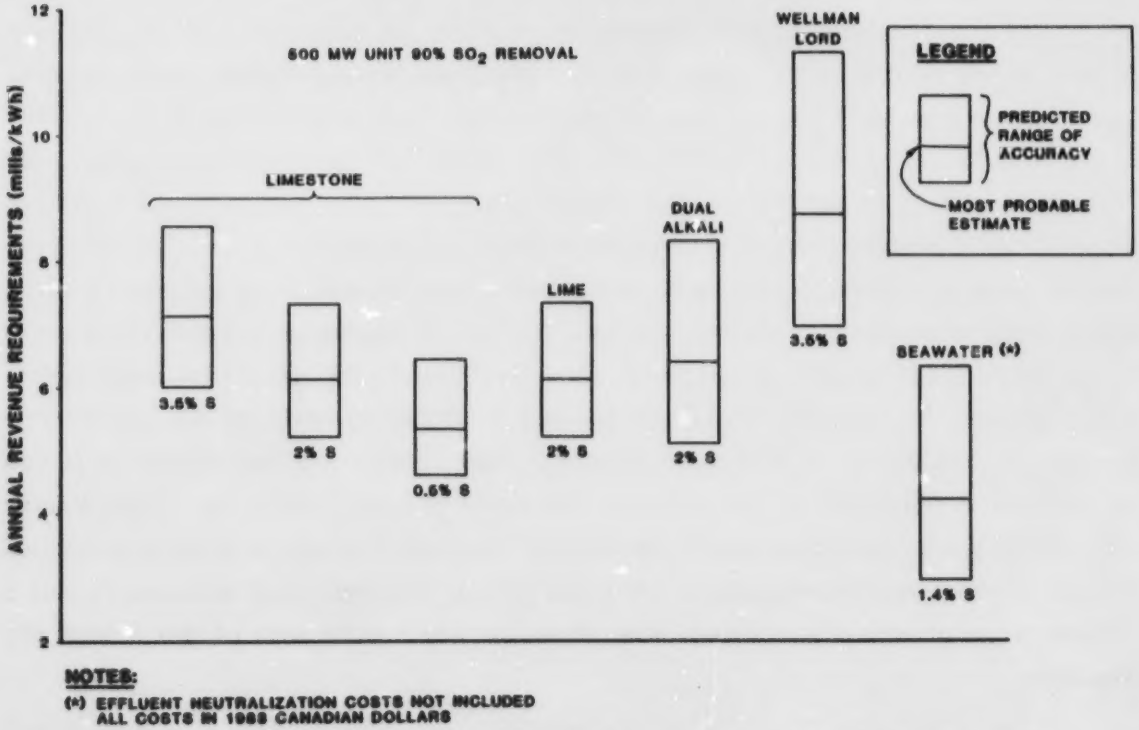


FIGURE 35 ANNUAL REVENUE REQUIREMENTS FOR ALTERNATIVE FGD PROCESSES REPORTED BY NATO (59)

The PEDCo data was based on actual reported costs for operating FGD systems in the U.S. utility industry. PEDCo normalized the costs reported for different facilities by deleting non-FGD-related costs and adding estimated costs for missing components in an attempt to standardize the scope of the facilities. However, no attempt was made to account for other sources of variance in cost such as unit size, sulphur content of coal, sulphur dioxide removal, new or retrofit installation, etc. A 30-year life is assumed for all new FGD systems and a 20-year life for all retrofit facilities. Capital costs include all direct equipment, installation and site development costs plus indirect costs such as interest, engineering, legal expenses, taxes, insurance, start-up and shakedown allowance, spare parts, contingency, land, and working capital. Annual revenue requirements include all raw materials, utilities, operating labour, maintenance and repairs, waste disposal and/or by-product sales, depreciation, insurance, taxes, and interest. The PEDCo data was likewise adjusted to reflect first quarter 1983 Canadian costs. Where possible, the data were categorized both by FGD process and approximate range of coal sulphur content. Figures 36 and 37 present the adjusted PEDCo capital costs and annual revenue requirements, respectively.

The range of costs illustrated in these figures on either side of the mean reflects the magnitude of the standard deviation for the reported costs. If the costs are assumed to follow a normal distribution, approximately 68 percent of reported costs will fall within the range illustrated.

Both NATO and PEDCo data illustrate a trend of reduced capital and operating cost for the limestone process at lower sulphur content in the fuel. They also show the relatively high cost of the Wellman-Lord process compared to the throwaway alternatives, even with a cost credit for by-product sulphur. The data also illustrate that, within the accuracy of the estimates, few conclusions can be drawn with regard to the relative costs of the lime, limestone, dual alkali, and alkaline fly ash processes. Although the seawater scrubbing process appears to be significantly less expensive than the other options considered, the NATO costs data are for a system without lime addition for effluent neutralization. The provision of lime neutralization and forced oxidation of the wastewater would add to both the capital cost and annual revenue requirements.

More recent work by TVA suggests that the actual costs of FGD are significantly higher than those presented above (60). TVA and Bechtel have developed a computer model which projects the design and costs for full-size lime and limestone FGD systems based on test results at the U.S. EPA Alkali Scrubbing Test Facility at the TVA Shawnee Steam Plant. Table 34 summarizes some predicted costs generated by the model

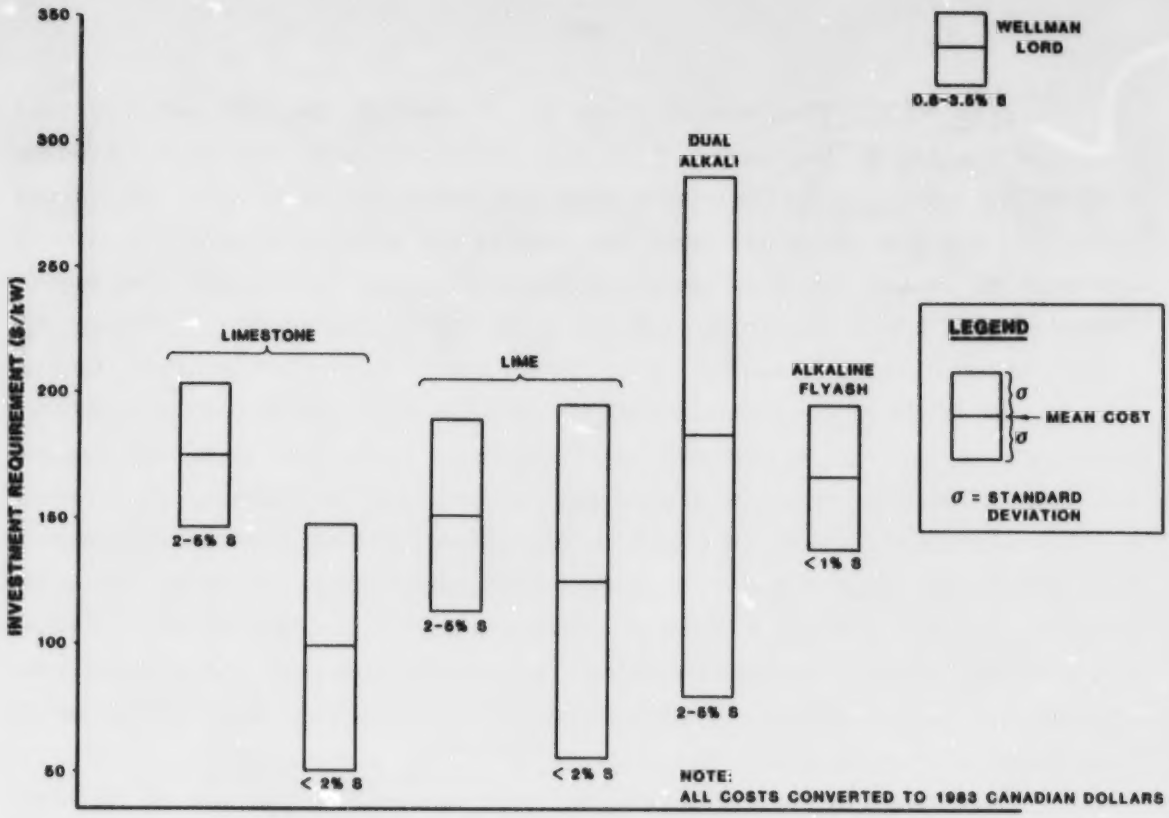


FIGURE 36 CAPITAL INVESTMENT REQUIREMENTS FOR ALTERNATIVE FGD PROCESSES REPORTED BY PEDCo (9)

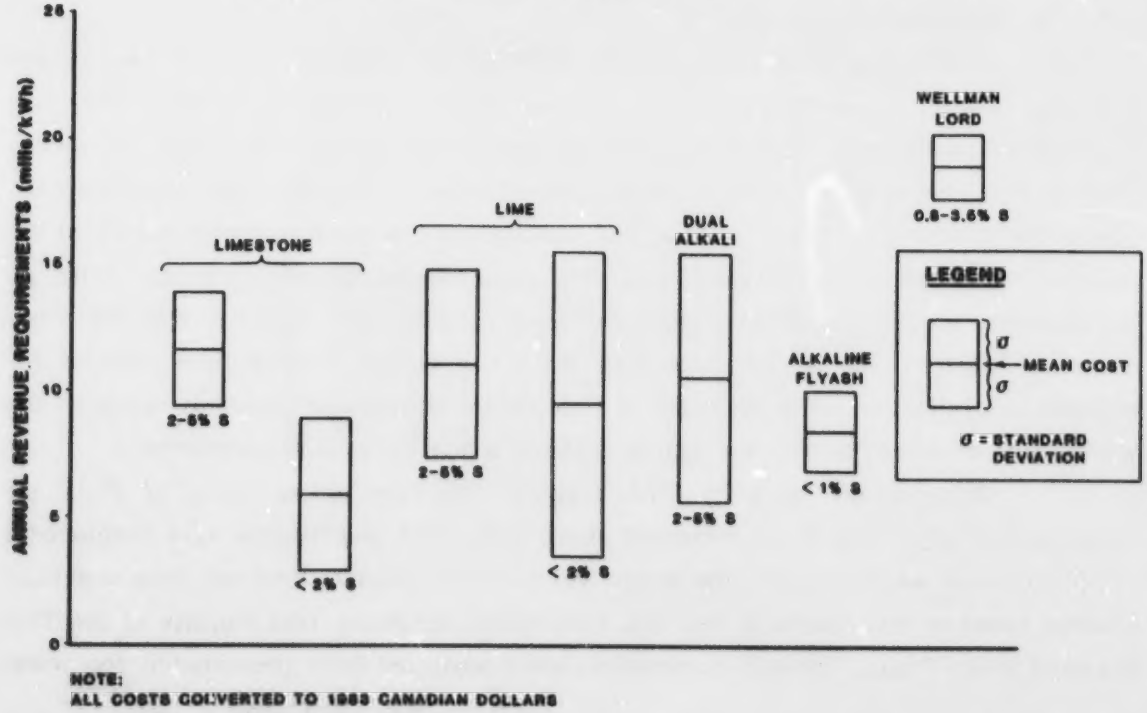


FIGURE 37 ANNUAL REVENUE REQUIREMENTS FOR ALTERNATIVE FGD PROCESSES BY PEDCo (9)

TABLE 34 LIMESTONE FGD SYSTEM COSTS[a] (60)

FGD System	Capital Cost[b] ($/kW)	First Year Revenue Requirement[c,d] (mills/kWh)	Levelized Annual Revenue Requirement[c,e] (mills/kWh)
Conventional Limestone - Natural oxidation - Pumping of sludge to a clay-lined disposal pond	225	10.5	14.4
Forced Oxidation Limestone - Gypsum dewatering - Trucking of gypsum to a clay-lined landfill	206	10.6	15.1
Advanced Limestone - Forced oxidation with adipic acid addition - Gypsum dewatering - Trucking of gypsum to a clay-lined landfill	186	9.4	13.3

a Based on a 500-MW boiler, burning 3.5% sulphur coal and operating 5500 hours/year. Accuracy of estimates = + 30%, - 15%.
b Reported in mid 1982 U.S. dollars.
c Reported in mid 1984 U.S. dollars.
d First year annual revenue requirements include raw materials, operating and overhead costs, and levelized capital charges.
e Levelized annual revenue requirements include factors to account for a 10%/year discount rate and a 6%/year inflation rate for operating costs over the life of the facility.

for a 500-MW limestone FGD system operating on 3.5 percent sulphur coal for three design options.

The recent TVA data estimate the capital cost and revenue requirements for the conventional limestone process to be significantly higher than those predicted by either the NATO or PEDCo data. It is believed that this is due largely to the higher costs associated with refinements in FGD system design in recent years. Both the PEDCo and NATO studies include costs for early systems. Reliability problems in many of these systems have been attributed to insufficient redundancy, use of inferior materials of construction, and other design-related problems. The TVA estimate is based on a system with one spare scrubber train, a spare feed preparation train and 50 percent emergency flue gas by-pass. This is more typical of the higher reliability FGD systems currently being installed.

The TVA model predicts a lower capital cost for the forced oxidation process than for the conventional limestone case. This is attributed to the high cost of pond construction associated with the latter, which more than offsets the costs savings associated with the elimination of forced oxidation equipment, thickener, filters, and landfill equipment. However, annual revenue requirements are slightly higher for the forced oxidation case due to the higher labour, electricity, diesel fuel, and overhead costs associated with the landfill operation. The model also predicts that the use of adipic acid or DBA enhanced limestone scrubbing results in about a 10 percent reduction in capital cost and annual revenue requirements, compared to the forced oxidation case without additives.

8.2 Waste Management Costs

The Electric Power Research Institute's FGD sludge disposal manual (26) reports cost data for several general classifications of sludge treatment and disposal systems. Adjusting these costs to 1983 Canadian conditions indicates that the reported cost of simple sludge ponding generally falls between $10 to $20 per tonne of dry solids. Most reported costs for more elaborate processing, such as stabilization or fixation, with landfilling are between $25 to $35 per tonne of dry solids (26).

One source estimated the capital investment and annual revenue requirements for several waste treatment and disposal options for a 500-MW limestone FGD system, operating on a 3.5 percent sulphur, 16 percent ash coal (61). The cost data, adjusted to 1983 Canadian conditions, are presented in Table 35. These data indicate that the forced oxidation process is lowest in capital investment requirements, and lower than all processes but untreated ponding in terms of annual revenue requirements. They also suggest that sludge disposal costs represent approximately 20 percent to 40 percent of the total FGD system capital and annual costs. However, the costs presented in Table 35 either include the cost of electrostatic precipitators or assume the removal of fly ash in the scrubber system. Caution must be exercised in making comparisons between cost estimates for alternative systems with dissimilar components.

As described in the previous section, TVA has found the capital investment requirements for conventional limestone scrubbing with ponding of sludge to be higher than forced oxidation with landfilling (60). These higher capital investments are largely the result of the disposal area construction costs, which reportedly are five to six times higher for ponds than for landfills. The lower operating and maintenance costs for ponds

TABLE 35 ESTIMATED CAPITAL INVESTMENT AND ANNUAL REVENUE REQUIREMENTS FOR ALTERNATIVE LIMESTONE FGD WASTE DISPOSAL SYSTEMS[a] (61)

Process	Capital Investment		Annual Revenue Requirements	
	Disposal Only ($/kW)	Scrubbers & Disposal[b] ($/kW)	Disposal Only (mills/kWh)	Scrubbers & Disposal (mills/kWh)
Untreated Ponding	56.7	176.6	1.45	6.67
Dravo (Calcilox) Ponding	79.4	199.3	2.95	8.17
Dravo (Calcilox) Landfill	41.7	161.6	2.92	8.13
Sludge - Fly Ash Blending	59.9[c]	180.0	2.53	7.74
Sludge - Fly Ash-Lime Blending (CSI)	35.3	155.1	2.33	7.56
Forced Oxidation (Gypsum)	25.4[d]	145.3	1.83	7.04

a Basis - new 500-MW plant; 3.5% sulphur, 16% ash in coal; limestone FGD process; one mile (1.6 km) to disposal site; all costs scaled to 1983 Canadian conditions.
b Basic limestone FGD system cost is $120/kW.
c Electrostatic precipitator cost ($31.7/kW) included.
d Air oxidation cost ($7.6/kW) included.

result in slightly lower annual revenue requirements; however, the differences are small when compared with the total cost of FGD.

Although gypsum stacking has yet to be practised on a commercial scale with FGD waste, this waste management option promises to offer the lowest cost of all alternatives considered. The need for sludge dewatering equipment (i.e., thickeners and filters) is eliminated as the scrubber bleed stream can be pumped directly to the stack. Operating labour requirements are minimal, and disposal area design and preparation is similar to a landfill. One Canadian utility evaluating FGD waste disposal options concluded that the capital and life cycle costs for forced oxidation with wet gypsum stacking, forced oxidation with landfill disposal, and fixation of sulphite sludge with dry fly ash were in the approximate proportions of 1:2:3, respectively (62).

As treatment of FGD wastewaters is not widely practised in North America at present, little data currently exist concerning treatment methods or costs. The most cost-effective treatment technology will obviously depend heavily upon both the size and characteristics of the wastewater stream and the effluent standard to which it must be treated. For instance, if TDS reduction is required to meet a discharge standard,

treatment costs can be expected to be several times greater than the cost of meeting Environment Canada's proposed discharge criteria for pH, suspended solids, and metals. A lime treatment process would likely be adequate to meet these criteria, and capital and operating costs for such a system would typically run in the range of $400 to $1200 per m^3/d capacity and $0.10 to $0.35 per m^3, respectively, depending on raw water chemistry and plant size (63).

TDS reduction would necessitate use of relatively more expensive membrane or evaporation processes. In an evaluation of vapour compression evaporation for the treatment of blowdown from a forced oxidation process, capital and operating costs, adjusted to 1983 Canadian values, were estimated to run from $2500 to $4300 per m^3/d of capacity and from $1.20 to $1.75 per m^3 treated, respectively (46). Some of this cost may be recoverable by using the treated effluent as makeup to the demineralizers of the boiler feedwater treatment plant. As evaporation distillate would typically be of higher quality than most raw water sources, some savings in pretreatment costs may be realized.

FGD wastewater treatment, including dissolved solids removal, warrants a detailed evaluation of the alternative technologies available and their associated costs.

9 CONCLUSIONS AND RECOMMENDATIONS

9.1 Conclusions

The following conclusions can be drawn from this study:

1) The FGD processes that have been commercially demonstrated in North America and are considered applicable to Canadian conditions include lime, limestone, dual alkali, alkaline fly ash, spray dryer, and Wellman-Lord processes. It has also been found that:

   a. lime and limestone processes have been applied to the full spectrum of coal sulphur contents, whereas dual alkali has been more commonly used with medium to high sulphur coals (2 to 5 percent sulphur) requiring high sulphur dioxide removals;

   b. alkaline fly ash and spray dryer processes have been used primarily on low sulphur (less than 1 percent sulphur) western lignite and subbituminous coals, although spray drying is making inroads in higher sulphur coal applications; and,

   c. although the Wellman-Lord process has been commercially used on coals with sulphur contents as low as 0.8 percent, it will generally be more applicable to high sulphur fuels, where higher by-product recoveries will better off-set the high capital and operating costs associated with the process.

2) Potential environmental concerns associated with FGD processes and their impact on land use include:

   a. the large land areas required for disposal of the large amounts of solid waste produced;

   b. the structural instability of unstabilized sulphite sludges, which may prevent reclamation of the disposal site;

   c. the chemical oxygen demand, metals and other dissolved solids associated with process wastewaters and disposal site leachates which may contaminate surface and groundwaters; and

   d. the increased water requirements which FGD processes impose on the power station.

3) Laboratory analyses performed on process wastewaters, potential wastewaters (e.g., disposal pond return water), and runoff from solid waste disposal sites indicated the following:

   a. The characteristics of these streams generally fell within Environment Canada's proposed discharge criteria for chromium, copper, iron, nickel, and zinc concentrations, although pH and suspended solids criteria were frequently exceeded.

   b. Treatability studies indicated that lime treatment produced an effluent which met the proposed discharge criteria from one sample in which the criteria were exceeded for iron and nickel concentrations.

4) As evaporative losses in the flue gas represent the bulk of the water loss from conventional wet scrubbing processes, the following conclusions can be drawn:

   a. There is no significant difference in makeup water requirements for the lime, limestone, dual alkali and alkaline fly ash processes for a specific coal.

   b. Spray dry scrubbing has approximately a 20 to 25 percent lower makeup water requirement than the wet scrubbing options examined in this study, because it operates above the adiabatic saturation temperature of the flue gas.

   c. The once-through seawater scrubbing process requires much larger makeup water volumes than other processes to provide the alkalinity necessary for scrubbing and the dilution necessary to prevent the precipitation of reaction products.

5) Given the state-of-the-art of FGD process design, it is considered that:

   a. closed-loop operation of natural oxidation lime, limestone, alkaline fly ash and dual alkali processes is a practical and attainable objective;

   b. the feasibility of closed-loop forced oxidation processes has not yet been demonstrated, and technical obstacles exist, particularly regarding the corrosion problems associated with high TDS operation, which adversely affect closed-loop operation;

   c. processes operated with a prescrubber loop, such as the Wellman-Lord, will normally require a blowdown for the control of TDS concentrations, particularly chlorides, in this circuit; and,

   d. TDS removal processes, such as mechanical evaporation, may be required to attain the zero-discharge objective for FGD systems that would otherwise require a blowdown of process water.

6) For a given coal type, the following conclusions can be drawn regarding solid waste volumes:

   a. The largest volume of FGD waste is produced by the ponding of sulphite sludge.
   b. Lower waste volumes are produced by landfilling dewatered sludge.
   c. Lowest waste volumes for wet scrubbing systems are produced by landfilling dewatered sludge from forced oxidation processes.
   d. Spray dry scrubbing produces lower volumes of FGD waste product than wet processes.

7) Current North American trends appear to be away from simple ponding of untreated sulphite sludges produced by wet scrubbers operating in a natural oxidation mode, in favour of:

   a. stabilization or fixation of sulphite sludges with fly ash (and lime) with disposal by landfilling;
   b. forced oxidation with disposal of gypsum by landfill, wet stacking, or utilization of by-product; and
   c. dry scrubbing with disposal of waste solids by landfill, following the addition of moisture for curing.

8) FGD systems have historically operated with a significantly lower reliability than other power generation equipment. Design improvements in newer systems have increased reliability considerably. These have included:

   a. the selection of more corrosion and erosion resistant materials of construction;
   b. increasing equipment redundancy; and
   c. providing a high degree of equipment winterization.

   These modifications have resulted in higher capital costs for new FGD systems.

9) Within the accuracy of the cost data compiled in this generic study:

   a. significant differences could not be detected between the wet scrubbing, throwaway processes considered;
   b. Wellman-Lord, the only regenerable process examined, appears to have significantly higher capital and operating costs, even when a full cost credit for by-product sulphur is included;

c. forced oxidation with landfill of waste solids appears to offer some reduction in capital cost but may have marginally higher annual revenue requirements than natural oxidation with ponding of untreated sludge; and

d. the use of adipic acid or DBA-enhanced limestone scrubbing may offer additional cost savings.

9.2 Recommendations

This review of water, wastewater, and solid waste management practices for FGD equipped coal-fuelled generating stations has identified several practices that will minimize the environmental effects associated with implementation of this technology in Canada. These are summarized below.

1) Dry Scrubbing - Consideration should be given to the selection of spray dryer FGD systems, particularly for low sulphur coal applications. The process offers simplicity of operation, the elimination of process wastewater, and a dry, manageable solid waste with good disposal properties as well as reuse potential.

2) Cascade Water Use - Wastewater streams from other processes within the station water system should be considered as potential sources of makeup to the FGD system. For example, discharge from auxiliary coolers or condensers could be used to minimize water use and the resulting entrainment of aquatic organisms.

3) Solid Waste Conditioning - Solid waste should be conditioned to produce a relatively stable product in terms of physical and chemical properties prior to disposal. This will normally entail either the use of forced oxidation to produce calcium sulphate with reuse potential, or fixation or stabilization of predominantly sulphite sludges and dry scrubbing wastes.

4) Solid Waste Disposal - Solid waste disposal areas should be designed in accordance with the following principles:

   a. Provision should be made to reclaim the site upon abandonment. Only physically and chemically stable material should be placed in the disposal area and, in the case of ponds, provision should be made to dewater the contents after interim storage of the sludge.

   b. Disposal areas should be designed to minimize contact between waste solids and water. Low permeability containment should be provided to prevent infiltration of groundwater, and to restrict seepage of leachate. Sites should be reclaimed in a modular fashion to minimize the amount of material exposed to the effects of precipitaiton and thereby limit runoff.

c. The disposal area should provide containment for all process flows and precipitation (infiltration and runoff) that it receives.

As these objectives are more readily met with dry disposal practices, landfilling is environmentally preferred to wet disposal methods, particularly in areas of net precipitation.

5) Wastewater Management - The following features should be incorporated in the wastewater management system:

a. sufficient surge capacity within the FGD system to accommodate normal process upsets and equipment drainings, without a wastewater discharge, and

b. flow measurement and sampling facilities to monitor all wastewater discharges. Groundwater quality in the vicinity of the solid waste disposal site should also be monitored on a routine basis.

Compliance with environmental guidelines may require that no process wastewater be discharged under normal operating conditions. While this has been accomplished with spray dryer and natural oxidation processes, further work is recommended to investigate the technical and cost implications of closed-loop operation with forced oxidation sytems.

6) Waste Utilization - Prospects for FGD waste utilization should be considered. The use of by-product gypsum from forced oxidation limestone processes appears to be the most economically feasible option at this time, although the technical and cost implications require further investigation.

Environment Canada is currently formulating more quantitative recommended waste management practices pertaining to the design of FGD systems. Recommended practices being considered by Environment Canada relate to six specific areas, as follows:

a) process design of FGD systems (closed process water loop);

b) physical stabilization of FGD solid waste (unconfined compressive strength of minimum 200 kPa);

c) modular development and reclamation of solid waste disposal sites;

d) containment of surface runoff from waste disposal sites (for a 100-year, 24-hour precipitation event);

e) seepage control to groundwaters (the equivalent of 1 m, 5×10^{-7} cm/s permeability liner between waste and aquifer);

f) wastewater effluent criteria (pH 6.5 - 9.5, Fe - 1.0 mg/L, Cr - 0.5 mg/L, Ca - 0.5 mg/L, Ni - 0.5 mg/L, Zn - 0.5 mg/L, TSS - 25 mg/L).

REFERENCES

1. Environment Canada, Clean Air Act, Thermal Power Generation Emissions -National Guidelines for New Stationary Sources, Canada Gazette, Part I, Extra No. 27, Vol. 115, Ottawa, April 25, 1981.

2. CH2M HILL Canada Ltd., "Development of Recommended Water, Wastewater, and Solid Waste Management Practices for FGD Equipped Coal-Fuelled Power Generating Stations - Phase One Report", prepared for Environmental Protection Service, Environment Canada, March 1982.

3. Dearborn Environmental Consulting Services, "Characterization and Treatability of Drainage Samples from Coal Piles at Canadian Steam Electric Stations", for Environment Canada, Report No. EPS 3-WP-82-4, Ottawa, 1982.

4. Shawinigan Engineering Western Limited, "Combustion of Canadian Coal and Lignite in Generating Station Boilers", prepared for the Canadian Electrical Association, Research Program 77-53, Calgary, October 1978.

5. T.E. Tibbetts, "Evaluation of Canadian Commercial Coals: Saskatchewan, Alberta and British Columbia - 1975", CANMET Report 76-41, Ottawa, August, 1976.

6. T.E. Tibbetts, and W.J. Montgomery, "Evaluation of Canadian Commercial Coals: Nova Scotia and New Brunswick - 1975", CANMET Report 76-40, Ottawa, August, 1976.

7. Dearborn Environmental Consulting Services, "Feasibility of Recirculating Bottom Ash Transport Water in Coal Fuelled Power Generating Stations", for Environment Canada, Report No. EPS 3-WP-83-1, Ottawa, 1983.

8. Anon, Hydrologic Atlas of Canada, Catalogue No. En37-26/1978, Fisheries and Environment Canada, 1978.

9. PEDCO Environmental Inc., "EPA Utility FGD Survey", U.S. Environmental Protection Agency, EPA 600/7-81-012C, 1982.

10. J. Makansi, "SO_2 Control: Optimizing Today's Processes for Utility and Industrial Powerplants", Power, Vol. 126, No. 10, October 1982.

11. T. Devitt, R. Gerstle, L. Gibbs, S. Hartman and R. Klier, "Flue Gas Desulphurization System Capabilities for Coal-Fired Steam Generators. U.S. EPA 600/7-78-032b, prepared by PEDCO - Environmental Inc., Cincinnati, Ohio, March 1978.

12. R.L. Sugarek, and T.G. Sipes, "Controlling SO_2 Emissions From Coal-Fired Steam Electric Generators: Water Pollution Impact", Volume II, Technical Discussion, U.S. EPA 600/7-78-045b, prepared by Radian Corporation, Austin, Texas, March, 1978.

13. L.A. Midkiff, "SO_2 Removal: Where do we stand?", Power, November 1978, pp. S21-26.

14. J. Meyler, "Dry Flue Gas Scrubbing - A Technique for the 1980's", Combustion, February 1981.

15. Flakt A/S Norsk Viftefabrikk, Flakt-Hydro Process: SO_2 Removal by Seawater, Technical Bulletin.

16. Environment Canada, "Environmental Code of Practice for Steam-Electric Power Generation", Draft, 1982.

17. A.L. Pollock, A. Berndl, and M. Esche, "A Lime Based FGD Process Utilizing Forced Oxidation to Produce Stable Gypsum By-Product", presented at Ontario Industrial Waste Conference, Toronto, Ontario, 1980.

18. L.J. Muzio, et al, "Demonstration of SO_2 Removal on a 22 MW Coal-Fired Utility Boiler By Dry Injection of Nahcolite", prepared for EPRI Conference on Fabric Filter Technology for Coal-Fired Power Plants, Denver, Colorado, 1981.

19. Anon., "A Proposed Procedure for the Development of a Canadian Data Base on Waste Leachability", Draft, Wastewater Technology Centre, Environment Canada, Burlington, Ontario, May 1981.

20. Anon., "Methods for Chemical Analysis of Water and Wastes", EPA 625/6-74-003a, Environmental Monitoring and Support Laboratory, Environmental Research Centre, Cincinnati, Ohio, 1976.

21. M.A. Franson (ed.), Standard Methods for the Examination of Water and Wastewater. Fifteenth edition, American Public Health Association, Washington, D.C., 1980.

22. P.G. Finlay, Personal Communication. Environmental Protection Service, Environment Canada, March 1982.

23. Committee D-19 on Water. Annual Book of ASTM Standards (Part 31), American Society for Testing and Materials, Philadelphia, Pennsylvania, 1978.

24. B.L. Tracy, and F.A. Prantl, "Radiological Implications of Thermal Power Production", presented at the International Symposium on Health Impacts of Different Sources of Energy, June 1981.

25. Anon., "Summary Report, Sulphur Oxides Control Technology Series: Flue Gas Desulphurization, Wellman-Lord Process", Industrial Environmental Research Laboratory, U.S. EPA 625/8-79-001, 1979.

26. Michael Baker Jr. Inc., "FGD Sludge Disposal Manual", Second Edition, Electric Power Research Institute, EPRI CS-1515, 1980.

27. Radian Corporation, "Controlling SO_2 Emissions From Coal-Fired Steam-Electric Generators - Water Polution Impact". Volume II Technical Discussion, Industrial Environmental Research Laboratory, U.S. EPA-600/7-78-045b, 1978.

28. PEDCo Environmental Inc., "Lime FGD Systems Data Book", Electric Power Research Institute and U.S. Environmental Protection Agency, EPRI FP-1030, 1979.

29. W. Ellison and E. Kutemeyer, "Commercial Forced-Oxidation FGD Technology, Input to J. Makansi for publication in Power February 1983.

30. G.T. Rochelle, "Buffer Additives for Limestone Scrubbing: A Review of R&D Results", presented at the EPA/EPRI Symposium on Flue Gas Desulphurization, Hollywood, Florida, 1982.

31. D.A. Burbank and S.C. Wang, "Adipic-Acid Enhanced Lime/Limestone Test Results at the EPA Alkali Scrubbing Test Facility - Capsule Report", U.S. Environmental Protection Agency 625/2-82-029, 1982.

32. J.C. Chang and J.H. Dempsey, "Pilot Plant Evaluation of By-Product Dibasic Acids as Buffer Additives for Limestone Flue Gas Desulphurization Systems", presented at the EPA/EPRI Symposium on Flue Gas Desulphurization, Hollywood, Florida, 1982.

33. Personal Communication, L. Luckevich, Ontario Research Foundation, 1983.

34. J.T. Yeh, R.J. Demski, D.F. Gyorke, and J.I. Joubert, "Experimental Evaluation of Spray Dryer Flue Gas Desulphurization for Use With Eastern U.S. Coals", Proceedings of the EPA/EPRI Symposium on Flue Gas Desulphurization, Hollywood, Florida, 1982.

35. W. Ellison and S.A. Lefton, "FGDS Reliability: What's Being Done to Achieve It", Power, May 1982.

36. Envirotech Corporation, "Sludge Dewatering For FGD Products", Electric Power Research Institute, EPRI FP-937, 1979.

37. Radian Corporation, "Chemical/Physical Stability of Flue Gas Cleaning Wastes", Electric Power Research Institute, EPRI FP-671, 1979.

38. R. Van Ness, A. Plumley, N. Mohn, C. Ullrich, and D. Hagerty, "Project Summary - Pilot Field Studies of FGD Waste Disposal at Louisville Gas and Electric", Industrial Environmental Research Laboratory, U.S. EPA 600/S7-81-155, 1981.

39. L.M. Pruce, "Evaluating the Newest Disposal Options For Scrubber Sludge", Power, May 1981.

40. C.M. Thompson and R.G. Rhudy, "Characteristics of Waste Products From Dry Scrubbing Systems", presented at EPA/EPRI Symposium on Flue Gas Desulphurization, Hollywood, Florida, 1982.

41. J.R. Donnelly, R.P. Ellis, and W.C. Webster, "Dry Flue Gas Desulphurization End-Product Disposal - Riverside Demonstration Facility Experience", EPA/EPRI Symposium on Flue Gas Desulphurization, Hollywood, Florida, May 1982.

42. J.C. Buschmann, E.L. Rasmussen, and S.M. Kaplan, "Disposal of Wastes From Dry SO_2 Removal Processes", presented at Joint Power Generating Conference, Phoenix, Arizona, 1980.

43. Personal Communication, Dr. W. Wells, Tennessee Valley Authority, Chattanooga, Tennessee, March 1983.

44. Ardaman & Associates, Inc., "Evaluation of Chiyoda Thoroughbred 121 FGD Process and Gypsum Stacking; Volume 3: Testing the Feasibility of Stacking FGD Gypsum", Electric Power Research Institute Report CS-1579, 1980.

45. Borgwardt, R., "Combined Flue Gas Desulphurization and Water Treatment in Coal-Fired Power Plants", Environmental Science and Technology, 14 (3), March 1980.

46. Southern Company Services, Inc., "Evaluation of Three 20 MW Prototype Flue Gas Desulphurization Processes, Electric Power Research Institute Report FP-713 Volume 2, 1978.

47. Personal Communication, D. Stonehouse, Minerals, Energy, Mines and Resources Canada, February 1982.

48. Energy, Mines and Resources Canada, Canadian Minerals Yearbook, Mineral Report 28, Ottawa, 1978.

49. Anon., "Gloomy Outlook for Natural Gas Sales", Oilweek, January 17, 1983.

50. Anon., "Sulphur Oxides Control in Japan", U.S. Interagency Energy/Environment Research and Development Program, EPA-600/9-79-043, November 1979.

51. J.W. Pressler, "Gypsum in 1981", Mineral Industry Surveys Preliminary Annual Report, U.S. Bureau of Mines, December 1981.

52. D.H. Stonehouse, Gypsum and Anhydrite - 1979, Energy Mines and Resources Canada, Ottawa, 1980.

53. F. Wirsching, H. Hamm, and R. Mueller, "Recycling of Desulfogypsum", United Nations Economic Commission for Europe, Third Seminar on Desulphurization of Fuels and Combustion Gases. Salzberg, Austria, 18-22 May, 1982.

54. Personal Communication, L. Luckevich, Ontario Research Foundation, 1983.

55. Goodwin, R.W., "Resource Recovery from Flue Gas Desulphurization Systems", Journal of the Air Pollution Control Association, 32 (9), September 1982.

56. Personal Communications, Dr. R. Bruce, Westroc Industries Limited, Mississauga, March 1982.

57. Winkler, P.W. et al., "Marketing Flue Gas Cleaning Process By-Products", presented at APCA Specialty Conference on: Waste Treatment and Disposal Aspects: Combustion and Air Pollution Control Processes, Charlotte, NC, Feb. 9-11, 1981.

58. Battelle Columbus Laboratories, "Analysis of Variations in Costs of FGD Systems", Electric Power Research Institute, 1978.

59. Tennessee Valley Authority, "Flue Gas Desulphurization Pilot Study; Phase II - Applicability Study", prepared for NATO Committee on the Challenges of Modern Society, 1980.

60. Burnett, T.A., Torstrick, R.L., and Sudhoff, F.A., "Economic Evaluation and Comparison o' Alternativ- Limestone Scrubbing Options", presented at the EPA/EPRI Symposium on Flue Gas D-sulphurization, Hollywood, Florida, 1982.

61. Barrier, J.W. et al., "E-onomics of D sposal of L - e/L - estone Scrubbing Wastes", U.S. Environmental Protection Agency, EPA 600/7-79-069.

62. Letter to P.G. Finlay, Power Generation and Energy Division, Environmental Protectio- Service, Environment Canada, from R.E. Waters, Ontario Hydro, December 1, 1982.

63. U.S. Environmental Protection Agency, "Innovative and Alternative T-chnology Assessment Manual", EPA-430/9-78-009, 1978.

APPENDIX A
SITE VISIT REPORTS

APPENDIX A - SITE VISIT REPORTS

Northern States Power Company
Sherburne County Generating Station (Becker, Minnesota)

Station/Unit Data

Units Scrubbed - No. 1 and No. 2
Generating Capacity - Unit 1, 740 MW (gross)
- Unit 2, 740 MW (gross)
Average Load Factor - 60-75%
Boiler Type - Pulverized Coal

Fuel Data

Coal Source - Montana Subbituminous (Colstrip and Sarpy Creek)
Average Heat Value - 20.3 MJ/kg
Average Ash Content - 9-10%
Average Moisture Content - 24-26%
Average Sulphur Content - 0.8-0.9%
CaO Content of Ash - variable from 2% to 20%

Particulate Removal. Particulates are removed in a venturi prescrubber (part of the FGD system) to meet an emission limit of 37 ng/J established by the state regulatory agency.

FGD System

Process - Limestone Alkaline Fly Ash
Supplier - Combustion Engineering
Installation - FGD systems installed new on both units
Unit No. 1 start-up: May 1976
Unit No. 2 start up: April 1977
SO_2 Removal Efficiency - Designed for minimum of 50 percent removal (emission limit is 413 ng/J on a 90-day rolling average and 516 ng/J for a 3-hour period).
- Actual SO_2 removal usually runs at 70-75%.
Absorbers - 12 countercurrent spray towers per unit.
- Designed with one spare module per unit.
- Actually operate absorbers about 20% above design gas flow rate and maintain two spare modules per unit.

Inlet Gas Flow - Design: 90-100 m^3/s per absorber
- Actual: 110-120 m^3/s per absorber

Inlet Gas Temperature - 93°C (low load)
- 177°C (high load)

Outlet Gas Temperature - 82°C
- Permit requires a minimum stack temperature of 71°C
- Reheat provided by an in-line hot water fin tube

Gas Pressure Drop - Average - 520 mm WG
Maximum - 635 mm WG

Reagent Consumption - 135-160 tonnes/day of limestone

Process Description. The Sherburne County Generating Station FGD system is illustrated schematically in Figure A-1. Flue gas from the air preheaters passes through a venturi prescrubber, where slurry is sprayed cocurrent to the gas flow for particulate removal. The flue gas then travels upwards through the absorber section of the scrubber, where slurry is sprayed countercurrent to the gas flow for SO_2 removal. It then passes through mist eliminators and over a hot water, finned-tube reheater, prior to discharge to the outlet duct.

The scrubbing slurry is maintained at a pH of approximately 5.0 to 5.5 by the addition of limestone slurry. Limestone is prepared by wet grinding in a ball mill using recycled water from the bottom ash transport system. Site runoff and FGD system drains are also added as makeup to this system.

The suspended solids concentration in the scrubber loop is maintained at approximately 10% by sending a bleed stream to a thickener. Thickener overflow is returned to the scrubber via a recirculation tank, while the underflow slurry is pumped to the fly ash pond at a solids concentration of 20-40%. Overflow from the fly ash pond plus cooling tower blowdown and Mississippi River water makeup are returned to the scrubber via the recirculation tank.

The water system was originally designed to operate open-loop with discharge to the river. However, as a result of design modifications, it has been operated closed-loop since the spring of 1978. Experience has indicated that when ash pond water is not recycled to the system, SO_2 removal, limestone utilization, and oxidation rate all decrease. The reason for this phenomenon is not understood.

Air is sparged into the reaction tank beneath the absorber to oxidize sulphite to sulphate. Near 100% oxidation is reportedly achieved when operating on ash pond recycle water.

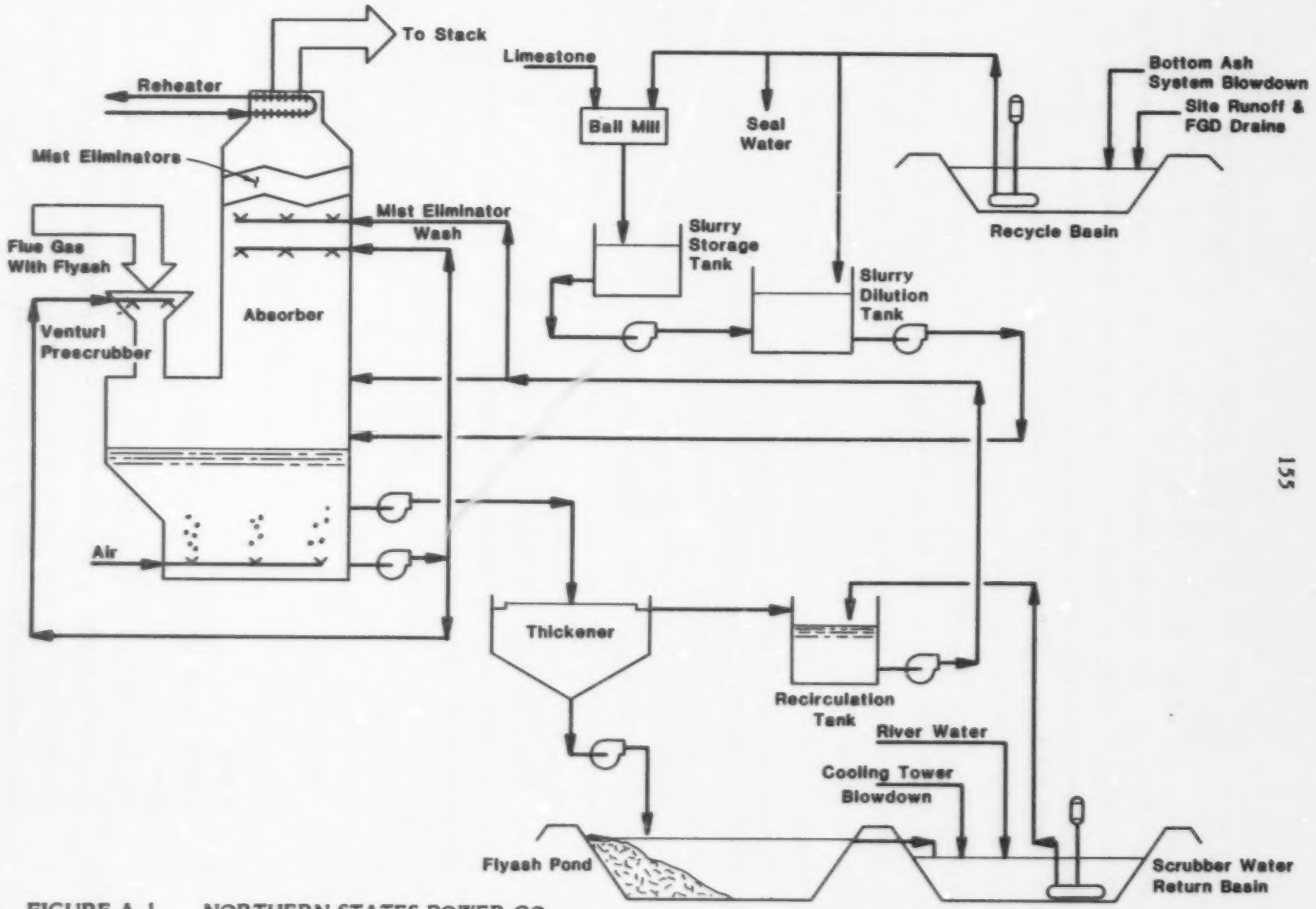


FIGURE A-1 NORTHERN STATES POWER CO.
SHERBURNE COUNTY GENERATING STATION
LIMESTONE ALKALINE FLY ASH FGD SYSTEM

Waste Disposal. The fly ash pond is the ultimate disposal site for the waste fly ash/scrubber sludge solids. The pond is approximately 15 m deep and has a capacity of 3 800 000 m^3, designed for a service life of 7 to 9 years. It is lined with native clay to minimize seepage. Monitoring wells are installed around the perimeter of the pond and groundwater is monitored monthly for parameters such as calcium, sulphate, and boron. The settled solids are reasonably stable, and reclamation and revegetation of the site is planned upon retirement of the pond.

Problems and Solutions. The FGD system has encountered numerous problems since start-up necessitating a major commitment by the operating staff to resolve these and achieve a 96 percent availability of the system in 1981. Some of the major problems are summarized below:

1) Control - The original design was for a highly automated system. However, due to numerous problems with instrumentation (particularly nuclear density meters) and with control valve failures, the system is now controlled in a semi-automatic/semi-manual mode. For instance, suspended solids determinations are now made daily in the laboratory and are used to manually control the bleed rate. Only three automatic control loops are used: slurry feed based on pH, reheat based on stack temperature, and recycle from the scrubber water return basin based on liquid level in the recirculation tank.
2) Marble Beds - The absorbers were initially equipped with a marble bed contacting section. However, due to problems with plugging, excessive gas-side pressure drop, and erosion of the perforated plate supporting the bed, they were converted to spray towers in 1980 and now use larger spray nozzles and higher L/G ratios than originally designed.
3) Materials - Materials of construction have been changed in many locations due to failure. These include:
   - External Slurry Piping - Original rubber-lined carbon steel has been converted to FRP.
   - Internal Spray Headers - Original FRP has been converted to 304 stainless steel.
   - Mist Eliminators - Original fibreglass is being converted to stainless steel due to fracturing caused by frequent washing with high pressure fire hoses.
   - Reaction Tank - Original carbon steel reaction tank floors have been replaced with concrete due to erosion and corrosion problems.

- Scrubber Lining - Numerous failures of the Flake lining have been encountered, which have been attributed to faulty application.

4) Plugging/Scaling - Scaling and "mud" deposition on scrubber internals is an ongoing problem necessitating a high degree of maintenance. A 12-man crew is responsible for manually cleaning mist eliminators, reheat tube bundles, strainers, etc. which are particularly prone to these problems. Approximately eight modules are taken off-line and cleaned daily in this manner.

With the exception of the thickeners, the FGD system is completely housed, which station personnel felt had contributed significantly to their ability to operate and maintain the system.

Cost. The total reported capital cost of the FGD system was approximately $69 000 000 U.S. in 1976/77. The waste disposal system for fly ash and scrubber sludge represented approximately $9 500 000 U.S. of the total cost.

Annual operating and maintenance cost data were not available. However, the FGD system has a staff of 51 people dedicated to its operation and maintenance; it has an energy consumption approximately equivalent to 40 MWe; and reagent costs for limestone represent approximately $1 000 000 U.S. per year.

Minnkota Power Cooperative Inc.
Milton R. Young Generating Station (Center, North Dakota)

Station/Unit Data

Units Scrubbed - No. 2 only
Generating Capacity - Unit 2, 440 MW (gross)
Load Factor - 73-84%
Boiler Type - Cyclone-fired

Fuel Data

Coal Source - Mine-mouth operation (Lignite)
Average Heat Value - 15 MJ/kg
Average Ash Content - 8%
Average Moisture Content - 37%
Average Sulphur Content - 0.4-0.9%

Particulate Removal. Approximately 30 percent of the coal ash leaves the cyclone-fired boiler as fly ash. This is captured in a cold-side electrostatic precipitator, designed for 99.6% removal efficiency, and fed to a fly ash silo. Approximately 2/3 of the fly ash is used as reagent to the FGD system. The remainder is hauled dry to mine disposal.

FGD System

Process - Lime Alkaline Fly Ash
Supplier - Combustion Equipment Associates
Installation - Designed during Unit 2 construction (Retrofit)
Unit No. 2 start-up: May 1977
FGD system start-up: June 1978
SO_2 Removal Efficiency - Design: Approximately 40% removal to meet 516 ng/J emission limit.
- Actual: 40-50% (achieve 80-90% removal on half of flue gas flow).
Absorbers - 2 spray towers with 50% capacity each.
- Only 1/2 of flue gas is scrubbed; therefore, one tower is spared.
- Dimensions - 12 m diameter x 36 m high.
Reagent Consumption - Lime is only fed when insufficient alkalinity is available in fly ash to control pH. At time of visit, no supplementary lime addition had been required for previous eight months.

Process Description. The Milton R. Young Generating Station scrubber flow diagram is shown in Figure A-2. Flue gas from the ESP's enters the absorber and travels upward through the dual-loop countercurrent spray tower. In the lower loop, which is operated at about 25% solids concentration, scrubbing slurry is circulated from the reaction tank in the absorber bottom to spray nozzles in the upper zones of the tower. A bleed stream from this loop is sent to a blowdown sump to control solids accumulation, and is ultimately pumped to a sludge pond for solids dewatering. Decant from the pond is returned to the system and is used to slurry fly ash, which is added to the lower loop for pH control. In the event insufficient fly ash is available to control pH in the 4.5 to 5.0 range, supplementary lime is slaked with fresh water and added to the lower loop.

In the upper loop, the mist eliminators are washed using recirculating wash water. The two loops are separated by a wash tray which captures the mist eliminator wash water. Suspended solids accumulation in this loop is controlled by sending a bleed stream to a clarifier. The settled solids are sent to a sludge pond, while the clarified overflow is retained to the upper loop. A portion of the clarifier overflow is used as wash tray underspray and thus becomes part of the lower loop. Its volume is made up with fresh water in the upper loop in an attempt to maintain a dissolved solids value below the saturation concentration of scaling salts in the wash water.

The FGD system operates closed-loop. Lake water is used as makeup to the system, although plant personnel indicated that they hope to eventually feed the scrubber and bottom ash systems with other plant wastewaters.

Waste Disposal. As originally designed and constructed, waste sludge was to be dewatered by gravity thickening followed by vacuum filtration. However, this system was abandoned in favour of sludge ponds due to operational problems. It was found that, to dewater the sludge adequately on the vacuum filter, the thickener had to be operated with such high solids concentrations in the underflow that plugging resulted. At solids concentrations low enough to avoid thickener problems, the sludge would not dewater sufficiently to facilitate solids handling. As a result, two large sludge ponds were constructed in 1981 to accept blowdown from the scrubber system. The ponds were clay-lined to a permeability specification of 2.5×10^{-7} cm/s and five groundwater monitoring wells are located about their perimeter. The ponds are used alternately, removing settled material from one while the other is in service. Approximately 1 m of coarse bottom ash is placed in the bottom of each pond prior to putting it into service. This material acts as an underdrain and provides the support and traction required for the front end loading equipment used to

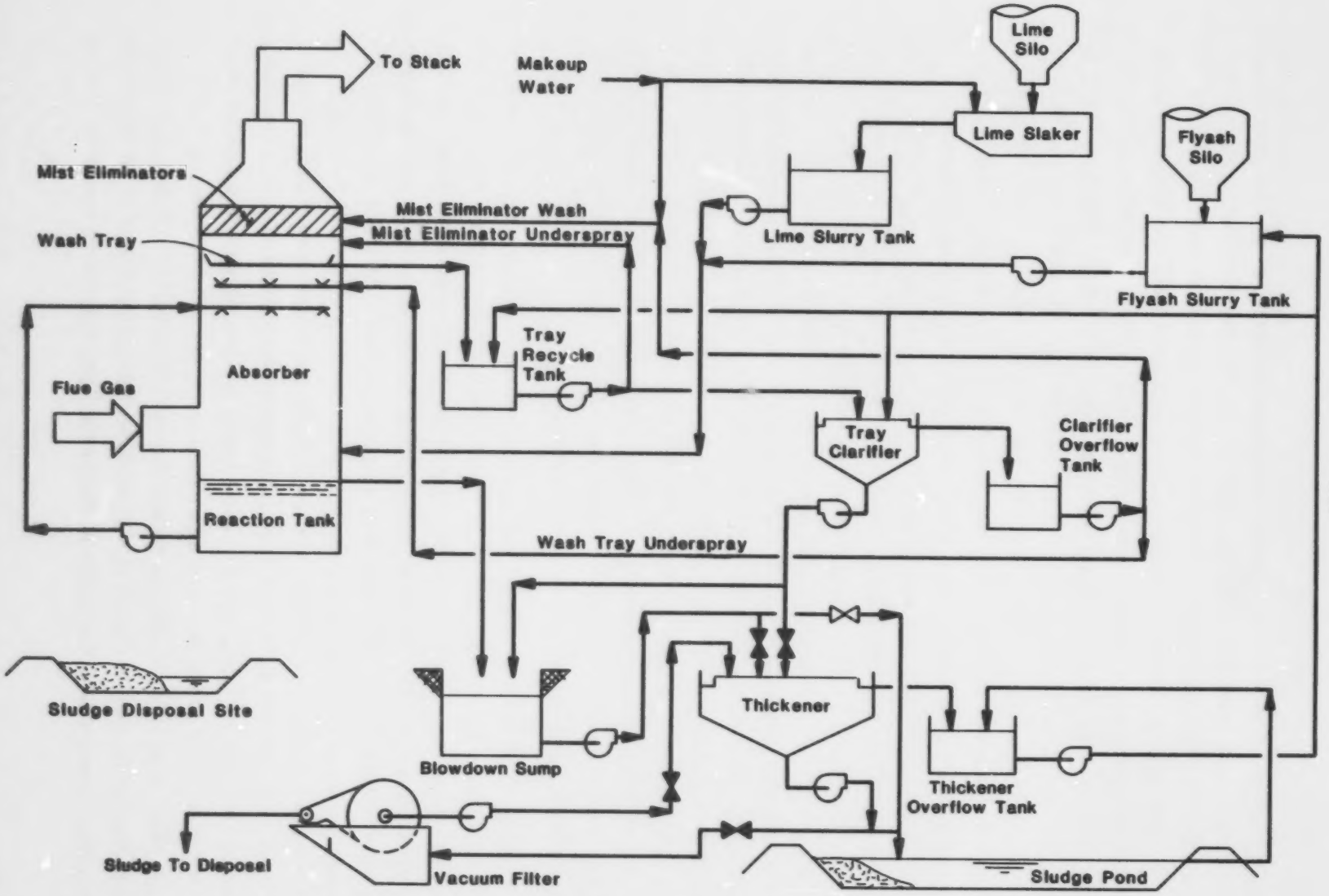


FIGURE A-2 MINNKOTA POWER COOPERATIVE INC.
MILTON R. YOUNG GENERATING STATION
LIME ALKALINE FLY ASH FGD SYSTEM

remove the material. The ponds are each equipped with decant structures which enable the water level in the ponds to be drawn down to the underdrain level, allowing occluded water to drain freely through the accumulated sludge prior to its removal.

At the time of the visit, sludge had been removed from one of the ponds for the first time. It reportedly was amenable to handling with front end loading equipment and trucks. The material is hauled to its ultimate disposal site in the coal mine.

Problems and Solutions. Besides the initial sludge dewatering problems described above, the FGD system has experienced the following operational problems:

1) Absorber Lining - Separation of the flakeglass lining from the absorber walls causes corrosion and leaks in the walls. This has been an ongoing problem and the lining is reapplied frequently when modules are taken out of service.
2) Demister Plugging - Scaling and plugging in the demisters has been a problem, particularly when operating with lime.
3) Control - Problems have been encountered with pH and density meters. In addition, system control is made difficult by an awkward mix of central control from the main station control room, and local control. Station personnel felt strongly that the scrubber system should have its own dedicated control room.
4) Piping - Erosion of piping has been an ongoing problem. Rubber-lined and unlined Schedule 80 carbon steel, FRP, and high density polyethylene have all been used with varying degrees of success.
5) Freezing - With the exception of chemical feed equipment, pumps, and vacuum filters, the FGD system is not housed. Equipment froze in the first few winters of operation. This was alleviated by the installation of heat tracing and insulation of exposed piping.
6) Scaling - A translucent, crystalline scale has formed on the pond water return line, significantly reducing the effective pipe diameter and thus the rate at which pond water can be recycled to the process.

Station personnel felt that to maintain high availability for the FGD system, spare scrubbing capacity is mandatory to enable one module to be maintained while the other is operating. They felt that at least one more absorber would be needed if the entire gas stream from Unit 2 required scrubbing.

Costs. The total reported capital cost of the FGD system was approximately $42 000 000 U.S. in 1978/79. The construction of the two sludge dewatering ponds cost an additional

$3 000 000 U.S. in 1981. Annual operating and maintenance costs were not available. However, the FGD system has a staff of four mechanics and one operator per shift and imposes an estimated energy penalty of approximately one percent of unit output.

Montana - Dakota Utilities Co.
Coyote Generating Station (Beulah, North Dakota)

Station/Unit Data

Units Scrubbed - 1
Generating Capacity - 440 MW (gross)
Load Factor - 75%
Boiler Type - Cyclone-fired

Fuel Data

Coal Source - Mine-mouth operation (Beulah Lignite)
Average Heat Value - 15.8 MJ/kg
Average Ash Content - 8-11%
Average Moisture Content - 34-36%
Average Sulphur Content - 0.6-1.1%
CaO Content of Ash - 13-14%

Particulate Removal. Particulates are removed in a fabric filter after spray drying of the flue gas. The baghouse is designed to meet a particulate emission limit of 43 ng/J.

FGD System

Process - Soda Ash Spray Drying
Supplier - Rockwell/Wheelabrator - Frye
Installation - New
- Start-up with station in April 1981

SO_2 Removal Efficiency - Design: To meet 516 ng/J emission limit (70% SO_2 removal with 75% reagent utilization).
- Actual: Has exceeded specifications in performance tests.

Absorbers - 4 spray towers at 14 m diameter by 4 m straight side.
- Designed for 33% capacity each at normal base load of 430 MW gross (405 MW net) with one unit spared (4th unit required at full load).
- Equipped with 3 rotary atomizers per absorber, rated at 112 kW with a nominal atomizing capacity of 1.5 L/s each.

Inlet Gas Flow - 520 N m^3/s (total design flow)
Inlet Gas Temperature - 140-150°C
Outlet Gas Temperature - 28-36°C above adiabatic saturation temperature
- Minimum outlet temperature = 71°C

Gas Pressure Drop - 310 mm WG across spray dryer
- 800 mm WG across fabric filters

Reagent Consumption
- 70 tonnes/day
- At 36°C approach to adiabatic saturation, 70% SO_2 removal can be achieved with 75% soda ash utilization.
- At 28°C approach to adiabatic saturation, 70% SO_2 removal can be achieved with greater than 90% utilization.

Process Description. The Coyote Station FGD system is illustrated schematically in Figure A-3. Flue gas enters the spray drying chamber where it is contacted with an atomized mist of soda ash scrubbing solution. Sufficiently low L/G ratios are used to ensure complete evaporation of the water within the liquor droplets. The fly ash and dried reaction products exit the bottom of the drier and are removed in a downstream fabric filter. The solids are transported from the baghouse hoppers to an ash disposal silo via a United Conveyors ash conveyor system. Moisture is added for dust suppression at the outlet of the silo and the solids are then hauled to a mined-out area for disposal.

Blowdown from the bottom ash pond is used as makeup water to the process for reagent dissolution, scrubbing solution dilution, and dust suppression. All plant wastewaters are directed to the bottom ash pond, except for cooling tower blowdown which is returned to the Missouri River, the source of supply for the station.

Primary system control is provided by two loops. Outlet gas temperature controls the flow of scrubbing liquor to the driers, and outlet SO_2 concentration controls the mix of saturated solution and ash pond water (i.e., reagent concentration).

Waste Disposal. Solids from the disposal silo are wetted and hauled to an abandoned area of the mine. The disposal site is lined with a native clay which was placed to meet a permeability specification of approximately 10^{-7} cm/s. Monitoring wells are installed around the disposal site and groundwater is sampled quarterly for analysis.

The material is placed and spread using trucks and scrapers. After placement, it reportedly cures to a very hard, impermeable texture with good load-bearing characteristics. The U.S. EPA extraction procedure has been run on the uncured material, and the laboratory leachates have apparently been found non-toxic under the Resource Conservation and Recovery Act (RCRA) regulations.

As the disposal area is located in a topographical depression, runoff from the surrounding mined out area collects in the site. This is pumped back to the station for use in the FGD system.

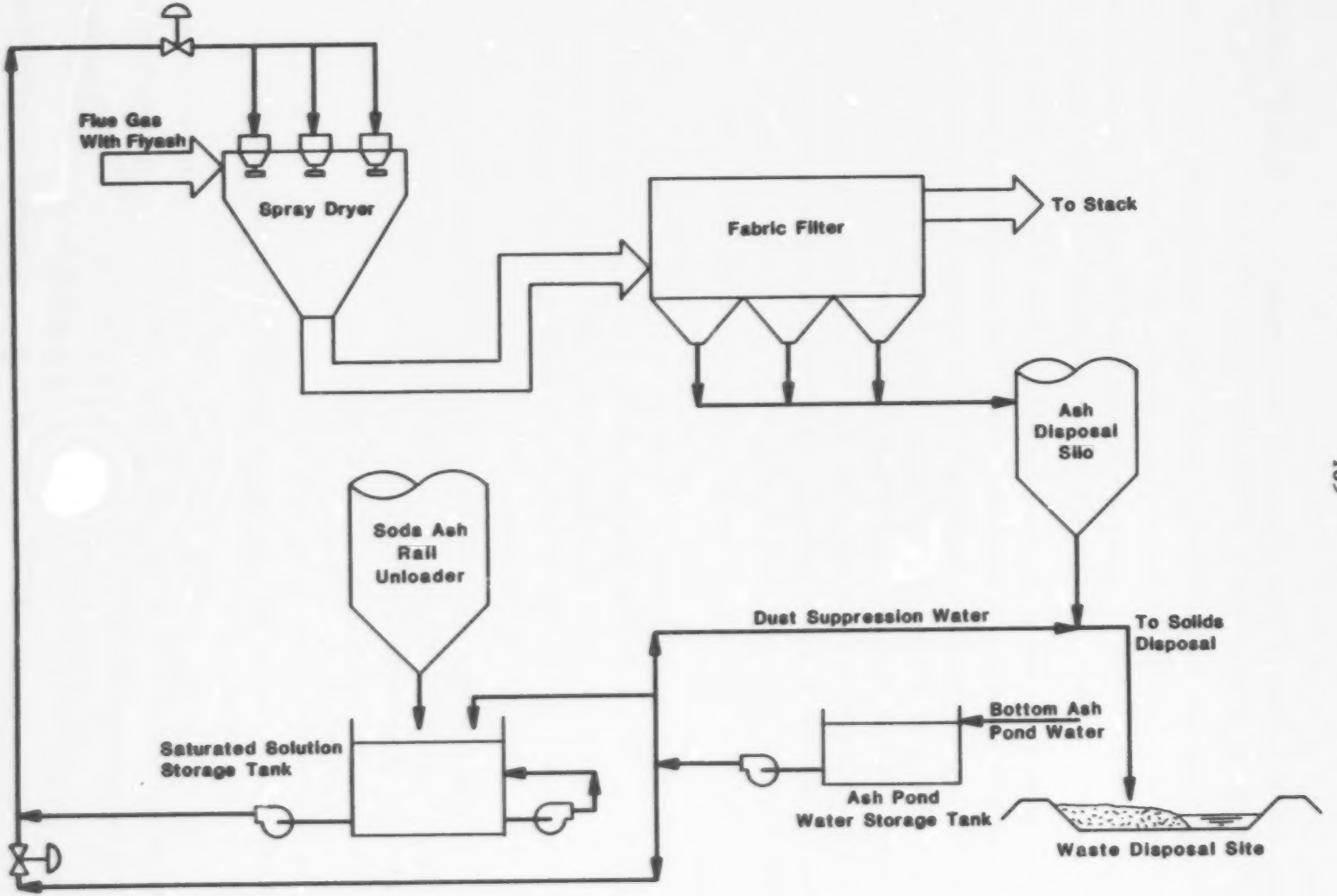


FIGURE A-3 MONTANA - DAKOTA UTILITIES CO.
COYOTE STATION
SODA ASH SPRAY DRYING FGD SYSTEM

Interest has been expressed by outside companies in use of the waste material in oil well cementing applications; tests are being carried out to determine its suitability.

Problems and Solutions. Station personnel reported that, although some load reduction has been experienced because all four scrubbing modules were not always available for service, the FGD system has not caused any forced boiler outages. Nonetheless, several shakedown problems had occurred in the first 18 months of service, as outlined below:

1) Ash Deposition - Deposition of ash on the scrubber walls has been attributed to poor atomization which leads to incomplete drying. The problem was alleviated by operating at higher outlet temperatures, but this lead to lower SO_2 removals and poorer reagent utilization. Several atomizer wheels have been tried and Rockwell-designed prototype atomizers have significantly improved performance.
2) Control - Sufficient pressure drop across the gas introducing vanes is required to ensure proper mixing and drying to keep droplets off the wall. Modules must be taken off-line as load drops to prevent the deposition of ash on the scrubber walls. Some difficulties have been experienced following transient loads with the FGD system.
3) Atomizer Drive System Failures - The output shaft on the atomizer rotates at 18 300 rpm. Several shaft and bearing failures in the gear box were experienced initially, necessitating frequent maintenance and repair. The drive mechanism and atomizer wheel have now been modified with a Rockwell-designed output shaft and wheel and tilt-pad bearings to improve alignment and balance.

In general, station personnel felt the system had performed well, particularly considering it is the first commercial scale system of its type. Performance has been very close to that predicted by early tests which were run using a 2-m diameter pilot unit. No corrosion or other materials problems have occurred.

Costs. The reported capital cost of the spray dryer/fabric filter system was approximately $38 000 000 U.S. in 1978, excluding concrete and electrical supply contracts. The system is completely housed.

The FGD system is run by two operators per shift. Reagent costs are estimated at approximately $4 000 000 U.S. per year. Waste disposal is handled on a subcontract basis at a cost of $0.87 U.S./tonne.

Northern States Power Company
Riverside Generating Station (Minneapolis, Minnesota)

Station/Unit Data

Units Scrubbed - No. 6 and No. 7
Generating Capacity - 110 MW (gross)
Boiler Type - Pulverized Coal

Fuel Data. In addition to its use for normal power production, the FGD system is being run as a demonstration unit and, as a result, different grades of coal are burned to evaluate its performance under varying conditions. Fuels burned have included a southern Illinois bituminous coal with 3.5% sulphur and an average heating value of 25 MJ/kg, a blend of 88% Colstrip subbituminous (sulphur content of 0.8%) and 12% petroleum coke (sulphur content of 4 - 4.5%), and Sarpy Creek (low sulphur subbituminous) coal in an 85-15% blend with petroleum coke, as well as straight 100%.

Particulate Removal. Particulates are removed in a fabric filter after spray drying of the flue gas. The baghouse reportedly produces an effluent gas stream with 2 to 3% opacity, well under the 20% opacity limit for the station.

FGD System

Process - Lime Spray Drying
Supplier - Joy Manufacturing/Niro Atomizer
Installation - New
- Spray Dryer start-up December 1980
- Baghouse start-up March 1981

SO_2 Removal Efficiency - Actual: 70-90% removal on subbituminous coal/petroleum coke blend (average 1.2% sulphur).

Absorber - 1 spray dryer at 14 m diameter by 11 m straight side (60° cone angled bottom).
- Compound gas disperser with 2 point discharge.
- Single rotary atomizer - motor rated at 522 kW
- gearbox, shaft, bearings, wheel, etc. rated at 600 kW

Inlet Gas Flow - Design: 275 m^3/s
- Typical: 205 m^3/s
- Maximum: 295 m^3/s

Inlet Gas Temperature - 138-177°C (with excursions to 193-204°C)

Outlet Gas Temperature - 62°C (10°C approach to adiabatic saturation temperature)

Gas Pressure Drop - Spray Dryer - 100 mm WG

Baghouse - less than 150 mm WG

Reagent Consumption - Less than 11 tonnes/day at typical operating conditions (stoiciometric ratio of 0.75 to 0.9). Higher and lower stoichiometric ratios have been used depending upon coal sulphur content and desired SO_2 removal.

Process Description. The Riverside Station FGD system is illustrated schematically in Figure A-4. Flue gas enters the spray drying chamber where it is contacted with an atomized mist of scrubbing slurry. A lime/fly ash slurry is used to cool the gas to approximately 10°C above its adiabatic saturation temperature. Approximately 10% to 20% of the fly ash and reaction products are disentrained from the gas stream within the dryer and are removed from the bottom of the chamber. The remaining 80% to 90% are carried out of the dryer with the scrubbed gas and are removed on a downstream fabric filter.

The system has the capability to recycle the solids (a Niro patented process) captured in the dryer and the baghouse by mixing them with water and fresh reagent and feeding the resultant slurry to the rotary atomizer. This apparently improves reagent utilization by reusing unreacted sorbent and utilizing the natural alkalinity of the fly ash.

Mississippi River water is used as makeup to the process for lime slaking, slurry dilution, and wetting of waste solids for dust suppression. Lime is slaked to 25% to 35% solids and diluted to approximately 20% percent. Recirculated solids are added to give a feed slurry strength of 35% to 45% solids, which is fed to the atomizer.

The system is equipped with two principal control loops. Outlet temperature controls slurry feed rate, although this is trimmed by a feed forward signal based on inlet gas flow and temperature. Lime feed rate is controlled by outlet SO_2 concentration or percent removal. This mix of water, lime, and recycle solids is predetermined and is set to maintain the desired solids concentration in the feed slurry.

Water Disposal. The reaction products from the process reportedly increase the total amount of solid waste produced by approximately 20% to 30% over the weight of fly ash alone. The reaction products apparently comprise about 70% calcium sulphite and 30% calcium sulphate. Water is added for dust suppression during handling, resulting in a final product with approximately 10% to 20% moisture.

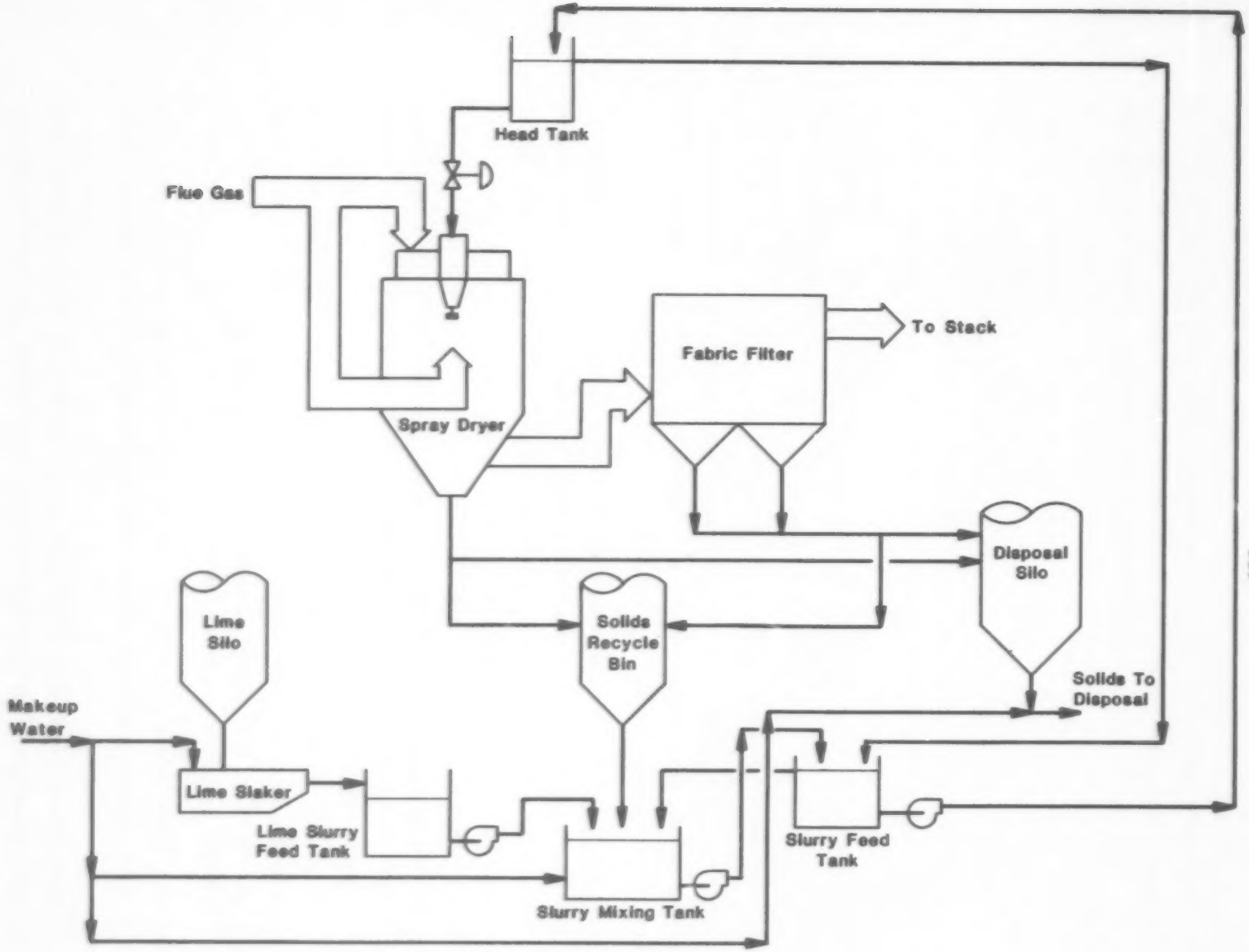


FIGURE A-4 NORTHERN STATES POWER CO.
RIVERSIDE GENERATING STATION
LIME SPRAY DRYING FGD SYSTEM

At the time of the visit, the FGD system was operated by Joy/Niro personnel, whereas Northern States Power Co. was responsible for waste disposal. No information was available regarding the disposal site, handling practices, or properties of the material after placement.

Several investigations have been conducted to examine the properties of the waste product and its reuse potential. Tests have indicated that the optimum moisture content for the material (i.e., moisture content which gives maximum dry density) varies with coal type, but typically ranges from 25 to 45%. Other physical characteristics of the material determined in the laboratory using waste samples at their optimum moisture content include the following:

Dry Density: 1.1 - 1.3 g/cm^3
28-Day Unconfined Compressive Strength: 1 - 10 MPa
Permeability: 0.5×10^{-6} - 8.4×10^{-6} cm/s

Leaching tests were also conducted using the U.S EPA extraction procedure which demonstrated the leachate would be classified as nontoxic under Resource Conservation and Recovery Act (RCRA) regulations.

Problems and Solutions. Joy/Niro personnel responsible for operation of the FGD system reported that relatively few operating problems have occurred. They indicated that in future systems they would recommend increasing redundancy, particularly with liquid handling equipment such as lime slakers, slurry tanks, pumps, etc. Reliable operation of the SO_2 analyzers had also apparently been difficult to achieve. A spray dryer bypass was also recommended to permit the absorber to be serviced while the boiler is operating. This would normally be accomplished by diverting the gas flow to a spare scrubbing module in a multiple-module system.

The stainless steel atomizer wheel used is equipped with silicon carbide inserts which reportedly give approximately 1000 to 2000 hours service per wear track. Four wear tracks per insert are available by rotating the insert one-quarter turn.

The FGD system is completed housed.

Costs. Because of the unique nature of the test facility at Riverside, Joy/Niro personnel indicated that the capital costs for the system were not representative and were not made available.

Operator requirements are estimated at two per shift for the FGD system plus one for the waste handling system up to approximately 400-500 MW capacity. In addition, two to four maintenance personnel, one shift per day, are recommended.

Power consumption for the lime preparation system was estimated at approximately 30% to 40% of the 522 kW requirement for the atomizer, while the baghouse is equipped with a 150 kW reverse air cleaning system.

Gulf Power Company
Scholz Generating Station (Chattahoochee, Florida)

Three prototype FGD systems, each of approximately 20 MW equivalent capacity, were pilot tested at the Scholz station between 1975 and 1977. These three processes were:

1) the Combustion Equipment Associates (CEA) dual alkali process,
2) the Foster-Wheeler/Bergbau-Forschung dry absorption process, and
3) the Chiyoda Thoroughbred 101 (CT-101) process.

The CT-101 process was subsequently modified by Chiyoda, which developed a new forced-oxidation limestone process, referred to as the Chiyoda Thoroughbred 121 (CT-121) process. This 23-MW prototype system was subsequently pilot tested at the Scholz station from October 1978 through May 1979.

As the CT-121 modification of the conventional limestone forced oxidation process was considered to be of interest from the Canadian perspective, a visit to the station was conducted to discuss the results of the pilot test with the plant manager. The results of the test programs are presented in detail in Electric Power Research Institute (EPRI) report FP-713-SY, "Evaluation of Three 20 MW Prototype Flue Gas Desulphurization Processes", Volumes 1, 2, and 3, and EPRI report CS-1579, "Evaluation of Chiyoda Thoroughbred 121 FGD Process and Gypsum Stacking", Volumes 1, 2, and 3 Addendum, and are briefly summarized below.

The CT-101 process, which has been widely used in Japan on oil-fired utility boilers, uses three principal reaction vessels: an absorber, an oxidizer, and a crystallizer. Flue gas that has been prescrubbed for removal of fly ash and chlorides flows through a packed absorption tower where SO_2 is absorbed, forming sulphurous acid (H_2SO_3). The sulphurous acid is oxidized to sulphuric acid (H_2SO_4) in an iron-catalyzed reaction in a separate oxidation vessel. The H_2SO_4 is then neutralized with limestone in a crystallizer vessel to produce gypsum. The gypsum crystals are separated from the mother liquor, while the supernatant is returned to the absorber.

The Scholz test represented the first application of the CT-101 process with combustion products from a coal-fired boiler. Over 25 months, the process demonstrated an availability of 81% and was successfully operated over a widely fluctuating range of inlet conditions with apparently little change in SO_2 removal efficiency, waste solids properties or limestone utilization. The system showed good resistance to short-term upsets and demonstrated little tendency towards scaling. In fact, no mist eliminator

washing was conducted during the entire test program of 10 600 hours, and yet no scaling was detected.

The CT-101 process could not be operated closed-loop, and initially had a wastewater discharge rate of approximately 0.6 to 1.3 L/s (0.03 - 0.06 L/s•MW), due primarily to excessive seal water requirements. Several modifications were made to replace makeup water with recycled scrubber liquor, which reduced the wastewater discharge by approximately 50%. A portion of the waste stream was treated by vapour compression evaporation, resulting in recovery of over 99% of the wastewater as high quality product containing less than 10 mg/L total dissolved solids.

The principal mechanical problem associated with the process involved the solid-bowl centrifuges used to dewater the gypsum crystal. Chiyoda has reportedly found basket-type centrifuges to be superior in this application in Japan. However, the solid-bowl type were used at Scholz because of size limitation. Other problems included failures of rubber-lined pumps, asbestos gas-side expansion joints, both epoxy and vinyl ester resin liners in the venturi prescrubber, and field-erected, hand-laid fibreglass joints in the oxidizer and associated piping.

At the conclusion of the CT-101 test program in March 1977, Chiyoda modified the demonstration equipment at Scholz to test their new CT-121 process. This system differs from CT-101 process in that SO_2 absorption, oxidation, and crystallization all occur within a single vessel, referred to as a jet bubbling reactor (JBR).

The process is illustrated schematically in Figure A-5. Flue gas is cooled and saturated in a venturi prescrubber prior to entering the JBR. At Scholz, a blend of approximately 25% fresh water and 75% pond recycle water was fed to the prescrubber to make-up for blowdown and evaporative losses. A continuous blowdown from this loop was necessary to prevent excess gypsum supersaturation due to evaporative concentration of the pond water. Prescrubber blowdown, at a pH between 2.0 and 2.5, was sent to either the JBR or gypsum tank for neutralization.

Gas from the prescrubber enters the JBR and is distributed to a number of sparger tubes. Pressure forces the gas beneath the slurry level in the reactor, causing the absorption of SO_2. Gas bubbles rise to the liquid surface in the JBR creating a froth layer in which SO_2 mass transfer is reportedly enhanced. Low superficial gas velocities (0.3-0.6 m/s) are used in the gas-liquid disengaging chamber at the top of the reactor to minimize the entrainment of slurry droplets in the gas stream. The gas leaving the JBR passes through mist eliminators before leaving the system through the outlet ductwork and stack.

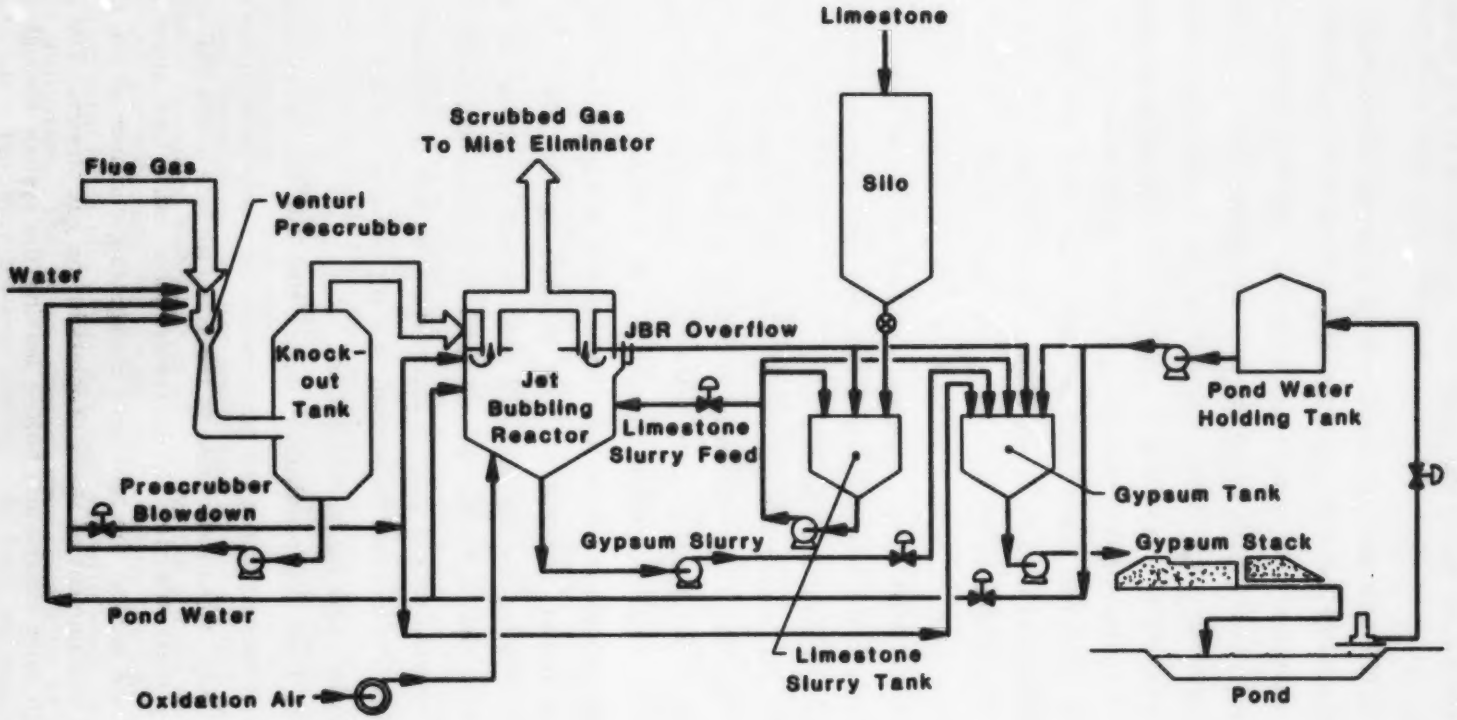


FIGURE A-5 GULF POWER COMPANY
SCHOLZ GENERATING STATION
CHIYODA THOROUGHBRED 121 FGD PROCESS

Slurried limestone is fed to the JBR to neutralize absorbed SO_2 and control pH. Compressed air is injected into the reactor to oxidize the reaction products to gypsum. A bleed stream of slurry is purged from the reactor to maintain the desired solids concentration in the JBR. At Scholz, this stream was sent to a gypsum tank, where limestone was added to raise the pH to about 6.0 for environmental reasons. The slurry was then diluted to about 5% to 10% solids with pond water or prescrubber blowdown to facilitate pumping to the disposal site. A gypsum stack was used for solids disposal and liquor was decanted from it to a separate pond. Pond water was returned and reused in the FGD system. Gypsum dewatering was also tested using a high-speed, rotary vacuum filter.

The test program conducted at Scholz involved both baseline tests conducted at Chiyoda-specified design conditions and tests at conditions outside of the design range. Process availability exceeded 99% for the duration of the six month test program. Process performance under design conditions is summarized below:

Coal Properties

Sulphur Content - 2.0%
Ash Content - 11.0%
Heating Value - 29.3 MJ/kg

Set Points

JBR Pressure Drop - 290 mm WG
JBR Overflow pH - 3.5
JBR Underflow Solids Concentration - 17%
Oxidation Air Rate - 0.58 Nm^3/s

Variables (average value)

Flue Gas Flow - 20 Nm^3/s
Inlet SO_2 Concentration - 1000-1200 ppm
SO_2 Removal - 92%
Limestone Utilization - 98%
SO_3 Oxidation to SO_4 - 100%

During operation under off-design conditions the effects on SO_2 removal of JBR pH, pressure drop, inlet SO_2 concentration O_2/SO_2 ratio, and limestone particle size were evaluated. No significant difference in SO_2 removal was observed when the

limestone particle size was changed from 90% less than 325 mesh to 90% less than 200 mesh. SO_2 removal decreased at O_2/SO_2 mole ratios of 4 or below, and with decreasing pressure drop and decreasing pH. SO_2 removal efficiency appeared to be relatively constant up to a gas phase SO_2 concentration of about 2500 ppm, beyond which it began to drop. Limestone utilization, measured around the JBR, remained about 98% under varying process conditions and the waste solids produced generally contained over 97% gypsum with essentially no sulphite solids.

The system operated with minimal scale deposition in the JBR or mist eliminators, although the mist eliminators were washed an average of only once a week for one minute with approximately 19 L/s of pond water.

Gypsum stacking was successful; the material settled to about 70% solids beneath the water surface. As the stack walls were raised by dredging the solids from the pond in the middle of the stack, the solids content of the gypsum at the top of the stack increased to over 90%. Vacuum filtering produced a filter cake typically containing about 85% solids directly from the JBR underflow stream, without prethickening. Some tests on waste product utilization demonstrated that the gypsum could be used in the manufacture of wallboard or the production of cement.

The system could not be operated closed-loop for most of the test period. During the winter, a large influx of rainfall to the gypsum stack and collection pond, coupled with a need to circulate fresh water through the supernatant return pipes for freeze protection, resulted in an estimated system overflow rate of approximately 0.5 L/s, although the average for the test was estimated at 0.2 L/s. A concentrated effort was made in the last three months of the test program to close the water loop. Pond water was used as seal water for two sets of pumps and the pond water/fresh water ratio was increased in the prescrubber. These measures reduced the fresh water requirement by a net amount of about 0.2 L/s.

In view of the success of the CT-121 test program, EPRI is reportedly arranging a 100-MW demonstration of the process.

Other Site Visit Reports

Six additional stations were visited in Phase One of this study. The station capacities, FGD processes used, and characteristics of coal burned at these stations are presented in Table 8. Detailed reports of the site visits to these stations are presented in Reference 2.

APPENDIX B
DETAILED ANALYTICAL RESULTS

TABLE B-1 CHARACTERIZATION OF FGD SYSTEM WATER STREAMS

Parameters	Milton R. Young Generating Station			Coyote Generating Station		Sherburne Generating Station			
	Decanted Blowdown Sump	Pond Return	Disposal Site Runoff	Disposal Site Runoff	Ash Pond Make Up Water	Thickener Overflow	Scrubber Make Up	Recycle Water	Ash Pond Overflow
pH	7.3	7.7	7.3	7.9	9.7	6.9	8.0	6.2	8.1
Specific Conductance (μmhos/cm)	13 000	3900	17 000	15 400	5000	11 000	8800	2400	8400
Total Solids	27 300	4700	45 970	35 540	5120	42 150	21 440	2890	20 670
Total Suspended Solids	36	17	126	40	136	945	108	29	103
Total Dissolved Solids	26 530	4690	45 100	34 330	5070	35 620	20 630	2880	20 590
Turbidity in N.T.U.	5.8	2.9	42	0.06	54	245	20	8.0	26
Chemical Oxygen Demand	12	13	12	72	25	204	150	15	158
"OH" Alkalinity as $CaCO_3$	0	0	0	210	0	0	8	0	10
Sulphate as SO_4	4420	460	6920	23 200	2480	22 100	15 700	1310	12 940
Sulphite as SO_3	< 1	< 1	< 1	< 1	< 1	< 1	< 1	< 1	< 1
Chloride	55.0	9.4	69.0	52.0	28.0	406	268	53.0	342
Fluoride	10.0	1.4	33.0	2.4	1.4	19.6	13.0	0.5	14.0
Orthophosphate as PO_4	0.15	< 0.05	0.20	0.25	1.3	1.0	0.35	0.10	0.15
Aluminum	< 0.05	< 0.05	3.92	0.90	2.89	50.6	1.53	< 0.05	2.57
Arsenic	0.038	< 0.001	0.040	0.310	0.130	0.009	0.005	< 0.001	< 0.001
Barium	< 0.005	0.008	< 0.005	0.044	0.029	< 0.005	0.061	0.028	0.045
Beryllium	< 0.0005	< 0.0005	< 0.0005	< 0.0005	< 0.0005	< 0.0005	< 0.0005	< 0.0005	< 0.0005
Boron	194	15.8	346	35.0	6.59	334	176	12.6	179
Cadmium	0.03	0.02	0.07	< 0.01	< 0.01	< 0.01	0.04	0.02	0.04
Calcium	420	396	442	132	40.5	483	480	470	479
Cobalt	< 0.05	< 0.05	< 0.05	< 0.05	< 0.05	< 0.05	< 0.05	< 0.05	< 0.05
Chromium	0.18	0.04	0.24	< 0.01	0.01	0.39	0.25	0.03	0.24
Copper	< 0.008	< 0.008	0.013	< 0.008	0.164	0.043	0.021	0.020	0.024
Iron	0.21	0.28	2.82	0.28	2.38	4.90	0.49	0.84	0.56
Lead	< 0.05	< 0.05	0.05	< 0.05	< 0.05	< 0.05	< 0.05	< 0.05	< 0.05
Lithium	1.5	0.2	2.4	0.4	0.1	4.0	2.4	0.3	2.4
Magnesium	1640	199	3200	20.0	22.7	5220	2580	115	2610
Manganese	1.26	0.24	1.00	< 0.01	0.04	49.7	14.1	0.38	13.7
Mercury (μg/L)	< 1	< 1	< 1	< 1	< 1	< 1	< 1	< 1	< 1
Molybdenum	0.8	0.5	0.4	0.3	< 0.3	2.5	2.0	0.4	1.9
Nickel	0.31	0.07	0.90	< 0.05	< 0.05	5100*	0.25	0.07	0.24
Phosphorus	1.4	0.8	1.9	< 0.6	1.4	2.1	2.0	0.9	2.0
Potassium	547	43	968	238	23	22	13	10	13
Selenium	0.040	< 0.001	0.001	0.115	0.016	0.26	0.120	< 0.001	0.121
Silica	6.89	1.16	31.9	4.26	4.16	33.2	< 0.05	11.8	< 0.05
Silver	< 0.005	< 0.005	0.053	< 0.005	< 0.005	< 0.005	< 0.005	< 0.005	< 0.008
Sodium	4020	547	6470	10 500	1520	593	356	85	356
Strontium	17.8	7.32	18.0	13.2	1.90	13.5	12.8	6.13	13.0
Thorium (μg/L)	< 3.5	< 3.5	< 3.5	< 3.5	< 3.5	6.2	< 3.5	< 3.5	< 3.5
Titanium	0.015	0.010	0.135	0.005	0.006	1.93	0.081	0.018	0.107
Uranium (μg/L)	53	11	13	7.7	14	53	33	< 3.0	< 3.5
Vanadium	0.084	0.013	0.102	0.358	0.093	0.193	0.134	0.020	0.131
Zinc	< 0.05	< 0.05	0.23	< 0.05	0.07	7.66	< 0.05	0.06	< 0.05
Zirconium	< 0.05	< 0.05	< 0.05	< 0.05	< 0.05	< 0.05	< 0.05	< 0.05	< 0.05

* Data Suspect - Duplicate analysis not available.
NOTE: All analyses reported in mg/L unless otherwise stated.

TABLE B-2 CHARACTERIZATION OF FGD SOLID WASTES

Parameter	Milton R. Young Station	Coyote Station		Sherburne County Station
		Dry Waste	Cured Waste	
Moisture Content (%)	35.2	0.2	31.0	29.1
Sulphate as SO_4	551 000	290 000	332 000	513 000
Sulphite as SO_3	< 1 000	35 000	51 000	12 500
Chloride	< 1 000	< 1 000	< 1 000	< 1 000
Phosphate as P_2O_5	24 000	2 000	24 000	< 1 000
Aluminum	57 000	51 500	37 800	47 800
Arsenic	197	103	9.3	10.7
Barium	8 760	8 690	2 060	6 730
Beryllium	5.2	1.5	1.5	0.5
Boron	1 730	1 830	1 290	392
Cadmium	< 1	< 1	< 1	< 1
Calcium	126 000	108 000	79 700	166 000
Cobalt	10	10	10	< 10
Chromium	60	50	30	20
Copper	124	133	104	32
Iron	64 700	40 600	31 900	13 800
Lead	190	75	55	20
Magnesium	16 300	26 700	20 900	9 810
Manganese	510	710	440	480
Mercury	2.86	0.73	0.72	0.75
Molybdenum	< 3	< 3	< 3	< 3
Nickel	630	30	30	20
Phosphorus	1 300	1 100	1 000	600
Potassium	8 000	6 000	4 000	2 000
Selenium	8	15	10	4
Silica	101 000	108 000	105 000	80 800
Silver	< 1	< 1	< 1	< 1
Sodium	27 000	115 000	160 000	6 000
Strontium	4 340	3 530	2 670	2 120
Thorium	7.9	12	8.7	11
Titanium	2 510	3 650	2 660	2 390
Uranium	8.7	8.0	5.9	5.2
Vanadium	93	111	72	26
Zinc	170	110	90	30
Zirconium	180	160	110	130

NOTE: All analyses reported in mg/kg (dry weight), unless otherwise stated.

TABLE B-3 CHARACTERIZATION OF LEACHING MEDIA AND LEACHATES OF FGD WASTES

Parameters	D.I. Medium	Acidic Medium	Milton R. Young G.S.				Coyote G.S.							
			D.I. Medium		Acidic Medium		Dry Sludge (Baghouse)				Cured Sludge			
							D.I. Medium		Acidic Medium		D.I. Medium		Acidic Medium	
			20:1	4:1	20:1	4:1	20:1	4:1	20:1	4:1	20:1	4:1	20:1	4:1
pH	7.5	4.4	7.5	7.2	4.6	5.5	11.5	12.0	8.6	11.1	11.6	12.1	7.4	9.8
Sp. Conductance (μmho/cm)	24	3000	3320	3900	6350	11 900	13 200	47 600	13 400	44 800	12 800	44 400	14 000	44 000
Total Dissolved Solids	7	3560	3700	4700	7860	16 100	15 350	78 760	23 220	80 800	16 500	72 200	23 720	76 000
Chemical Oxygen Demand	3	7580	3	2	6170	5500	118	1180	6940	7450	186	1200	6460	6930
"OH" Alkalinity as $CaCO_3$	0	0	0	0	0	0	380	940	0	950	420	2330	0	830
Sulphate as SO_4	< 1	< 1	2460	3210	2110	6090	9570	41 800	2980	42 400	6150	36 400	7020	35 000
Sulphite as SO_3	< 1	< 1	< 1	< 1	< 1	< 1	1.1	9600	3	6800	2300	12 900	< 1	12 400
Chloride	2.0	4.0	6.0	7.0	9.0	25.0	40.0	145	35.0	140	25.0	65.0	17.0	70.0
Fluoride	0.10	NA*	2.1	NA*	NA*	NA*	7.1	NA*	NA*	NA*	3.6	NA*	NA*	NA*
Orthophosphate as PO_4	< 0.05	< 0.05	< 0.05	< 0.05	0.2	< 0.05	0.2	2.7	1.3	2.1	0.6	4.5	0.3	2.6
Aluminum	< 0.01	< 0.01	< 0.01	< 0.01	33.7	5.42	46.8	69.9	< 0.01	93.0	41.4	137	< 0.01	41.6
Arsenic	< 0.005	< 0.005	< 0.005	< 0.005	< 0.005	< 0.005	0.220	5.80	0.190	1.170	0.560	3.700	0.070	1.24
Barium	< 0.005	< 0.005	0.025	0.021	0.019	0.009	0.023	< 0.005	< 0.005	< 0.005	0.031	0.018	0.034	0.015
Beryllium	< 0.0005	< 0.0005	< 0.0005	< 0.0005	0.0037	< 0.0005	< 0.0005	< 0.0005	< 0.0005	< 0.0005	< 0.0005	< 0.0005	< 0.0005	< 0.0005
Boron	< 0.004	< 0.004	10.9	36.6	21.8	72.7	22.6	46.6	37.6	92.2	19.7	71.9	29.3	97.8
Cadmium	< 0.01	< 0.01	< 0.01	< 0.01	< 0.01	< 0.01	< 0.01	< 0.01	< 0.01	< 0.01	< 0.01	< 0.01	< 0.01	< 0.01
Calcium	< 0.01	< 0.01	496	468	799	596	165	124	652	255	48.6	65.7	376	43.2
Cobalt	< 0.05	< 0.05	0.10	< 0.05	0.07	< 0.05	< 0.05	< 0.05	< 0.05	< 0.05	< 0.05	< 0.05	< 0.05	< 0.05
Chromium	< 0.01	< 0.01	< 0.01	< 0.01	0.02	< 0.01	0.05	0.06	< 0.01	< 0.01	0.01	0.02	< 0.01	< 0.01
Copper	< 0.008	< 0.008	0.012	0.011	0.098	0.020	< 0.008	< 0.008	< 0.008	< 0.008	< 0.008	< 0.008	< 0.008	< 0.008
Iron	< 0.01	< 0.01	0.06	0.09	4.64	0.25	0.04	< 0.01	0.10	< 0.01	< 0.01	< 0.01	0.02	< 0.01
Lead	< 0.05	< 0.05	< 0.05	< 0.05	< 0.05	< 0.05	< 0.05	< 0.05	< 0.05	< 0.05	< 0.05	< 0.05	< 0.05	< 0.05
Lithium	< 0.1	< 0.1	0.3	0.4	0.3	1.0	0.5	1.4	1.0	1.0	0.1	0.3	0.4	0.3
Magnesium	< 0.01	< 0.01	156	195	72.9	1080	0.04	0.09	276	0.19	0.15	0.14	220	3.35
Manganese	< 0.01	< 0.01	0.03	0.17	1.11	2.39	< 0.01	< 0.01	1.43	< 0.01	< 0.01	< 0.01	1.44	< 0.01
Mercury (μg/L)	< 1	< 1	< 1	< 1	< 1	< 1	< 1	15.5	< 1	< 1	< 1	21.0	< 1	< 1
Molybdenum	< 0.3	< 0.3	0.8	0.8	< 0.3	< 0.3	0.5	2.2	0.7	2.3	< 0.3	0.9	< 0.3	0.8
Nickel	< 0.05	< 0.05	0.09	0.09	0.14	0.28	< 0.05	< 0.05	0.13	< 0.05	< 0.05	< 0.05	0.07	< 0.05
Phosphorus	< 0.6	< 0.6	1.1	1.0	1.1	1.1	< 0.6	0.6	1.3	< 0.6	< 0.6	< 0.6	< 0.6	< 0.6
Potassium	< 1	< 1	30	75	44	158	106	516	115	509	50	227	56	235
Selenium	< 0.005	< 0.005	< 0.005	0.006	0.006	< 0.005	0.520	3.1	0.600	2.200	0.170	0.800	0.093	0.730
Silica	0.06	< 0.05	1.77	3.67	72.7	35.4	4.14	9.56	29.4	1.69	7.83	7.35	18.8	0.35
Silver	< 0.005	< 0.005	0.012	< 0.005	< 0.005	< 0.005	< 0.005	< 0.005	< 0.005	< 0.005	< 0.005	< 0.005	< 0.005	< 0.005
Sodium	< 1	909	240	469	1020	1950	4350	22 300	5190	23 300	4810	21 800	5830	21 700
Strontium	0.004	< 0.001	8.58	10.7	15.8	11.9	20.2	26.1	41.7	25.3	4.47	7.31	21.4	2.55
Thorium (μg/L)	< 3.5	< 3.5	< 3.5	< 3.5	< 3.5	< 3.5	< 3.5	< 3.5	< 3.5	< 3.5	< 3.5	< 3.5	< 3.5	< 3.5
Titanium	< 0.005	< 0.005	< 0.005	< 0.005	< 0.029	< 0.005	< 0.005	< 0.005	< 0.005	< 0.005	0.005	0.00	< 0.005	< 0.005
Uranium (μg/L)	< 3.0	< 3.0	< 3.0	< 3.0	43	29	< 10	< 3.0	33	28	< 3.5	< 12	45	95
Vanadium	< 0.005	< 0.005	0.023	0.016	< 0.005	0.021	0.498	1.67	0.446	0.971	0.474	1.34	0.056	0.483
Zinc	< 0.05	< 0.05	0.08	0.10	0.26	0.38	< 0.05	< 0.05	0.08	< 0.05	< 0.05	< 0.05	0.07	< 0.05
Zirconium	< 0.05	< 0.05	< 0.05	< 0.05	< 0.05	< 0.05	< 0.05	< 0.05	< 0.05	< 0.05	< 0.05	< 0.05	< 0.05	< 0.05

NOTE: All analyses in mg/L unless otherwise specified.
*NA Not available due to interference.
D.I. Medium - Deionized Water Medium

TABLE B-3 CHARACTERIZATION OF LEACHING MEDIA AND LEACHATES OF FGD WASTES (Cont'd)

	Sherburne G.S.			
	D.I. Medium		Acidic Medium	
Parameters	20:1	4:1	20:1	4:1
pH	9.4	9.1	4.8	6.8
Sp. Conductance (μmhos/cm)	2250	2360	6040	8300
Total Dissolved Solids	2530	2620	9200	11 500
Chemical Oxygen Demand	5	10	6300	5430
"OH" Alkalinity as $CaCO_3$	12	28	0	0
Sulphate as SO_4^{-2}	1580	1770	1740	1560
Sulphite as SO_3	< 1	< 1	< 1	< 1
Chloride	5.0	17.0	7.0	21.0
Fluoride	1.4	5.3	NA*	NA*
Orthophosphate as PO_4	< 0.05	< 0.05	1.5	0.05
Aluminum	1.42	2.08	107	< 0.01
Arsenic	0.008	< 0.005	< 0.005	< 0.005
Barium	0.125	0.096	0.042	< 0.005
Beryllium	< 0.0005	< 0.0005	< 0.0005	< 0.0005
Boron	5.86	20.8	11.4	40.6
Cadmium	< 0.01	< 0.01	< 0.01	< 0.01
Calcium	613	624	1.090	1 540
Cobalt	0.09	< 0.05	< 0.05	< 0.05
Chromium	0.03	0.13	0.07	0.05
Copper	0.010	< 0.008	0.082	0.011
Iron	0.02	0.04	3.61	0.11
Lead	< 0.05	< 0.05	< 0.05	< 0.05
Lithium	0.1	0.2	0.4	0.6
Magnesium	3.48	10.1	86.5	138
Manganese	< 0.01	< 0.01	5.01	8.55
Mercury (μg/L)	< 1	< 1	< 1	< 1
Molybdenum	< 0.3	0.3	< 0.3	0.5
Nickel	0.07	0.07	0.15	0.16
Phosphorus	0.9	0.7	2.3	1.4
Potassium	< 1	4	< 1	5
Selenium	0.019	0.022	0.013	0.008
Silica	0.60	0.34	92.4	20.0
Silver	< 0.005	< 0.005	< 0.005	0.010
Sodium	14	39	946	904
Strontium	5.75	10.2	12.6	22.8
Thorium (μg/L)	< 3.5	< 3.5	5.7	< 3.5
Titanium	< 0.005	0.005	0.807	< 0.005
Uranium (μg/L)	< 3.0	< 10	46	31
Vanadium	0.035	0.033	0.023	0.011
Zinc	0.05	< 0.05	0.20	0.23
Zirconium	< 0.05	< 0.05	0.08	< 0.05

NOTE: All analyses in mg/L unless otherwise specified.
*NA Not available due to interference.
D.I. Medium - Deionized Water Medium

TABLE B-4(a) GROSS ALPHA AND BETA ACTIVITY IN FGD SOLID WASTES AND WASTEWATERS FROM MILTON YOUNG, COYOTE AND SHERBURNE GENERATING STATIONS

		Activity					
		Solid Waste[a]		Wastewaters[b]			
				Gross Alpha		Gross Beta	
Station	Sample	Gross Alpha	Gross Beta	Solid[c]	Liquid[d]	Solid[c]	Total[e]
Milton Young	Sludge	1.5	1.1	-	-	-	-
	Blowdown Sump	-	-	<0.04	<0.04	0.22	21
	Sludge Runoff	-	-	<0.04	<0.04	0.37	0.37
	Pond Return Water	-	-	0.04	<0.04	0.04	0.37
Coyote	Dry Sludge	1.5	0.74	-	-	-	-
	Cured Sludge	1.1	0.74	-	-	-	-
	Sludge Runoff	-	-	<0.04	<0.04	0.04	5.5
	Makeup Water	-	-	0.04	<0.04	<0.04	0.04
Sherburne	Sludge	0.37	0.37	-	-	-	-
	Thickener Overflow	-	-	0.26	<0.04	0.74	1.1
	Scrubber Makeup	-	-	0.07	<0.04	0.26	0.37
	Recycle Water	-	-	<0.04	<0.04	0.04	0.37
	Sludge runoff	-	-	0.07	<0.04	0.22	0.37

a Reported in units of Bq/g on a dry weight basis.
b Reported in units of Bq/L, determined on 3-10 mL of unfiltered sample.
c Direct count of suspended solids filtered out of 100 mL of sample.
d By collection of radionuclides on $BaSO_4$ from filtrate alone.
e Total beta count using 3 - 10 mL of unfiltered sample.

TABLE B-4(b) RADIONUCLIDES IN WATER SAMPLES FROM MILTON YOUNG GENERATING STATION

Parameter	Analytical Method	Wastewaters			Leachates			
					Acidic Medium		D.I. Medium	
		Blowdown Sump	Disposal Site Runoff	Pond Overflow	4:1	20:1	4:1	20:1
Gross alpha	-	<6	<1	<1	<1	<1	<1	<1
Gross beta	-	25 ± 20	18 ± 9	<3	<3	<3	<3	<3
Gross gamma*	-	63	15	5	2.3	0.9	1	0.5
Th 228	Alpha Spectroscopy	0.80	0.25	0.07	0.07	<0.01	<0.01	0.06
Th 230	Alpha Spectroscopy	1.52	0.65	0.45	0.75	<0.01	0.16	0.70
Th 232	Alpha Spectroscopy	0.91	0.25	0.06	0.12	<0.01	0.01	0.05
Th 234	Alpha Spectroscopy	2.5	<0.5	<0.5	<0.5	<0.5	<0.5	<0.5
Ra 226	Alpha Spectroscopy	2.8	<0.2	<0.2	<0.02	0.3	<0.2	0.2
Ra 222	Alpha Spectroscopy	2.3	<0.1	<0.1	<0.1	<0.1	<0.1	<0.1
Pb 214	Alpha Spectroscopy	2.3	<0.1	<0.1	<0.1	<0.1	<0.1	<0.1
Bi 214	Alpha Spectroscopy	2.3	<0.1	<0.1	<0.1	<0.1	<0.1	<0.1
Th 228	Gamma Spectroscopy	0.12	<0.04	<0.04	<0.04	<0.04	<0.04	<0.04
Ra 220	Gamma Spectroscopy	0.12	<0.04	<0.04	<0.04	<0.04	<0.04	<0.04
Pb 212	Gamma Spectroscopy	0.12	<0.04	<0.04	<0.04	<0.04	<0.04	<0.04
Po 212	Gamma Spectroscopy	0.12	<0.04	<0.04	<0.04	<0.04	<0.04	<0.04

NOTE: All results expressed in Bq/L.
* Gross gamma 0 to 2 MeV.

TABLE B-5 QUALITY CONTROL ANALYSES FOR REPRODUCIBILITY OF ANALYTICAL METHODS AND LEACHING PROTOCOL

	Milton Young		Coyote						Sherburne	
	D.I. Medium Leachate 4:1 Liquid:Solid		Dry Waste Acidic Medium Leachate 4:1 Liquid:Solid		Cured Waste Acidic Medium Leachate 20:1 Liquid:Solid		Dry Waste		Acidic Medium Leachate 20:1 Liquid:Solid	
Parameter	Run 1	Run 2	Run 1	Run 2	Run 1	Run 2	Run 1	Run 2	Run 1	Run 2
Aluminum	< 0.01	< 0.01	93.0	104	41.4	38.6	52 500	51 500	107	99.8
Barium	0.021	0.026	< 0.005	0.087	0.031	0.168	8310	8690	0.042	0.049
Beryllium	< 0.0005	< 0.0005	< 0.0005	0.0005	< 0.0005	< 0.0005	1.7	1.5	< 0.0005	< 0.0005
Boron	36.6	45.1	92.2	80.9	19.7	20.4	1810	1830	11.4	11.7
Cadmium	0.03	0.02	< 0.01	0.01	< 0.01	< 0.01	< 0.01	< 0.01	0.03	0.03
Calcium	464	439	255	265	48.6	44.6	108 000	108 000	1090	1040
Cobalt	< 0.05	< 0.05	< 0.05	< 0.05	< 0.05	< 0.05	< 10	< 10	< 0.05	< 0.05
Chromium	< 0.01	< 0.01	< 0.01	0.03	0.01	0.02	40	50	0.07	0.06
Copper	0.011	< 0.008	< 0.008	0.008	< 0.008	< 0.008	132	133	0.082	0.030
Iron	0.09	0.12	< 0.01	< 0.01	< 0.01	< 0.01	41 700	40 600	3.61	2.70
Lead	< 0.05	< 0.05	< 0.05	< 0.05	< 0.05	< 0.05	75	75	< 0.05	< 0.05
Magnesium	195	148	0.19	0.19	0.15	0.68	27 000	26 700	86.5	122
Manganese	0.17	0.23	< 0.01	< 0.01	< 0.01	< 0.01	690	710	5.01	4.54
Molybdenum	0.8	0.7	2.3	2.3	< 0.3	< 0.3	< 0.3	< 0.3	< 0.3	< 0.3
Nickel	0.09	0.08	< 0.05	< 0.05	< 0.05	< 0.05	30	30	0.15	0.15
Phosphorus	1.0	0.9	< 0.6	0.7	< 0.6	< 0.6	1200	1100	2.3	2.2
Potassium	75	64	509	501	50	49	6000	6000	< 1	< 1
Silica	3.67	4.38	1.69	1.87	7.83	7.82	108 000	108 000	92.4	90.5
Silver	< 0.005	< 0.005	< 0.005	< 0.005	< 0.005	< 0.005	1	< 1	< 0.005	< 0.005
Sodium	449	328	23 300	23 600	4810	4650	115 000	115 000	946	855
Strontium	10.7	10.0	25.3	24.6	4.47	4.19	3510	3530	12.6	12.1
Titanium	< 0.005	< 0.005	< 0.005	< 0.005	< 0.005	< 0.005	3640	3650	0.807	0.562
Vanadium	0.016	< 0.005	0.971	1.01	0.494	0.488	99	111	0.023	0.012
Zinc	0.10	0.08	< 0.05	< 0.05	< 0.05	< 0.05	110	110	0.20	0.19
Zirconium	< 0.05	< 0.05	< 0.05	< 0.05	< 0.05	< 0.05	160	160	0.08	0.06

Note: All analyses in mg/L